

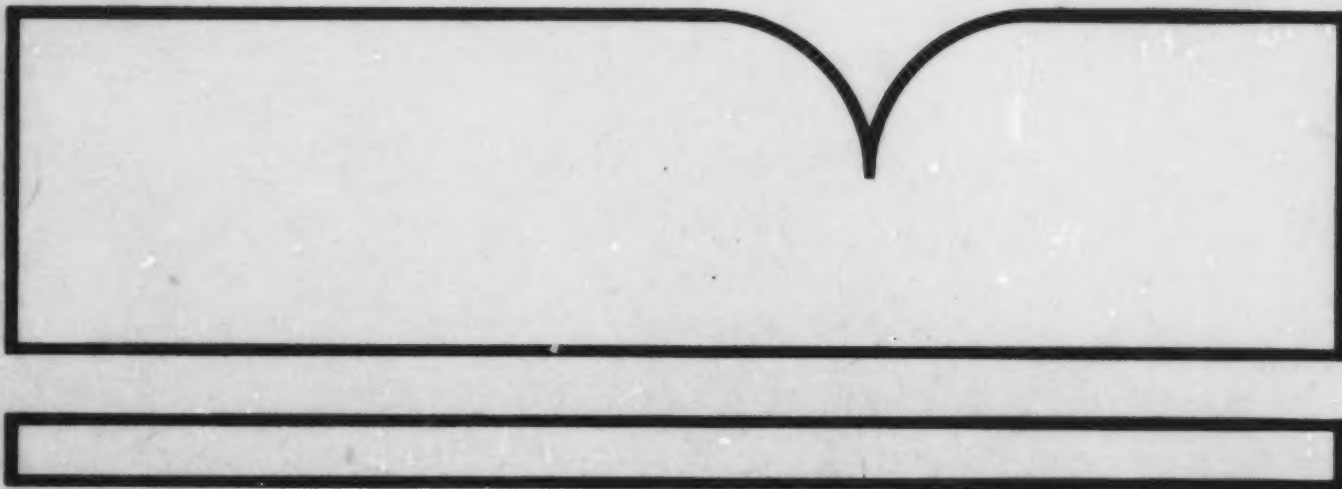
Organic Chemical Manufacturing. Volume 4
Combustion Control Devices

IT Enviroscience
Knoxville, TN

Prepared for

Environmental Protection Agency
Research Triangle Park, NC

Dec 80



U.S. Department of Commerce
National Technical Information Service

NTIS

PB81-220535

United States
Environmental Protection
Agency

Office of Air Quality
Planning and Standards
Research Triangle Park NC 27711

EPA-450/3-80-026
December 1980

Air



Organic Chemical Manufacturing Volume 4: Combustion Control Devices

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TECHNICAL REPORT DATA (Please read Instructions on the reverse before completing)		
1. REPORT NO. EPA-450/3-80-026	2.	3. REPORT'S ACCESSION NO. PMT 220535
4. TITLE AND SUBTITLE Organic Chemical Manufacturing Volume 4: Combustion Control Devices	5. REPORT DATE December 1980	
	6. PERFORMING ORGANIZATION CODE	
7. AUTHOR(S) J. W. Blackburn J. A. Key H. S. Basdekis V. Kalcevic	8. PERFORMING ORGANIZATION REPORT NO.	
9. PERFORMING ORGANIZATION NAME AND ADDRESS IT Envirosience, Inc. 9041 Executive Park Drive Suite 226 Knoxville, Tennessee 37923	10. PROGRAM ELEMENT NO.	
	11. CONTRACT/GRANT NO. 68-02-2577	
12. SPONSORING AGENCY NAME AND ADDRESS DAA for Air Quality Planning and Standards Office of Air, Noise, and Radiation U.S. Environmental Protection Agency Research Triangle Park, North Carolina 27711	13. TYPE OF REPORT AND PERIOD COVERED Final	
	14. SPONSORING AGENCY CODE EPA/200/04	
15. SUPPLEMENTARY NOTES		
Reproduced from best available copy.		
16. ABSTRACT EPA is developing new source performance standards under Section 111 of the Clean Air Act and national emission standards for hazardous air pollutants under Section 112 for volatile organic compound emissions (VOC) from organic chemical manufacturing facilities. In support of this effort, data were gathered on chemical processing routes, VOC emissions, control techniques, control costs, and environmental impacts resulting from control. These data have been analyzed and assimilated into the ten volumes comprising this report. This volume covers the following devices that can be used to control VOC emissions: thermal incinerators, catalytic incinerators, and flares.		
17. KEY WORDS AND DOCUMENT ANALYSIS		
a. DESCRIPTORS	b. IDENTIFIERS/OPEN ENDED TERMS	c. COSATI Field/Group
		13B
18. DISTRIBUTION STATEMENT Unlimited Distribution	19. SECURITY CLASS (This Report) Unclassified	21. NO. OF PAGES 354
	20. SECURITY CLASS (This page) Unclassified	22. PRICE

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EPA-450/3-80-026

**Organic Chemical Manufacturing
Volume 4: Combustion
Control Devices**

Emission Standards and Engineering Division

U.S. ENVIRONMENTAL PROTECTION AGENCY
Office of Air, Noise, and Radiation
Office of Air Quality Planning and Standards
Research Triangle Park, North Carolina 27711

December 1980

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INTRODUCTION

A. SOCHI PROGRAM

Concern over widespread violation of the national ambient air quality standard for ozone (formerly photochemical oxidants) and over the presence of a number of toxic and potentially toxic chemicals in the atmosphere led the Environmental Protection Agency to initiate standards development programs for the control of volatile organic compound (VOC) emissions. The program goals were to reduce emissions through three mechanisms: (1) publication of Control Techniques Guidelines to be used by state and local air pollution control agencies in developing and revising regulations for existing sources; (2) promulgation of New Source Performance Standards according to Section 111(b) of the Clean Air Act; and (3) promulgation, as appropriate, of National Emission Standards for Hazardous Air Pollutants under Section 112 of the Clean Air Act. Most of the effort was to center on the development of New Source Performance Standards.

One program in particular focused on the synthetic organic chemical manufacturing industry (SOCHI), that is, the industry consisting of those facilities primarily producing basic and intermediate organics from petroleum feedstock materials. The potentially broad program scope was reduced by concentrating on the production of the nearly 400 higher volume, higher volatility chemicals estimated to account for a great majority of overall industry emissions. EPA anticipated developing generic regulations, applicable across chemical and process lines, since it would be practically impossible to develop separate regulations for 400 chemicals within a reasonable time frame.

To handle the considerable task of gathering, assembling, and analyzing data to support standards for this diverse and complex industry, EPA solicited the technical assistance of IT Enviroscience, Inc., of Knoxville, Tennessee (EPA Contract No. 68-02-2577). IT Enviroscience was asked to investigate emissions and emission controls for a wide range of important organic chemicals. Their efforts focused on the four major chemical plant emission areas: process vents, storage tanks, fugitive sources, and secondary sources (i.e., liquid, solid, and aqueous waste treatment facilities that can emit VOC).

B. REPORTS

To develop reasonable support for regulations, IT Enviroscience gathered data on about 150 major chemicals and studied in-depth the manufacture of about 40 chemical products and product families. These chemicals were chosen considering their total VOC emissions from production, the potential toxicity of emissions, and to encompass the significant unit processes and operations used by the industry. From the in-depth studies and related investigations, IT Enviroscience prepared 53 individual reports that were assembled into 10 volumes. These ten volumes are listed below:

- Volume 1 : Study Summary
- Volume 2 : Process Sources
- Volume 3 : Storage, Fugitive, and Secondary Sources
- Volume 4 : Combustion Control Devices
- Volume 5 : Adsorption, Condensation, and Absorption Devices
- Volume 6-10: Selected Processes

Volumes 4 and 5 are dedicated to the evaluation of control devices used as add-on controls to reduce VOC emissions. These add-on controls are discussed generally in Volumes 2 and 3 as emission control options for the control of VOC emissions from generic sources. The use of these add-on controls in specific applications is demonstrated in the process studies covered in Volumes 6 through 10.

This volume covers the application of combustion devices as add-on VOC emission control devices. Separate reports are presented covering control device evaluations for thermal oxidation, special thermal oxidation requirements for VOC containing halogens and sulfur, catalytic oxidation, flares, and the use of emissions as fuels. These reports discuss the practical use of each control device, describe the systems, and discuss key design considerations. Data, tables, and curves are presented to enable preliminary cost and energy impacts to be determined for a wide range of potential applications. These control device evaluation reports were used to develop the cost effectiveness and energy impact determinations presented in the process reports of Volumes 6 through 10. The focus of these reports is on control of new sources rather than on existing sources in keeping with the main program objective of developing new source performance standards for the industry. The reports do not

outline regulations and are not intended for that purpose, but they do provide a data base for regulation development by the EPA.

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REPORT 1

CONTROL DEVICE EVALUATION
THERMAL OXIDATION

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July 1980

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ABBREVIATIONS AND CONVERSION FACTORS

EPA policy is to express all measurements used in agency documents in metric units. Listed below are the International System of Units (SI) abbreviations and conversion factors for this report.

<u>To Convert From</u>	<u>To</u>	<u>Multiply By</u>
Pascal (Pa)	Atmosphere (760 mm Hg)	9.870×10^{-6}
Joule (J)	British thermal unit (Btu)	9.480×10^{-4}
Degree Celsius ($^{\circ}\text{C}$)	Degree Fahrenheit ($^{\circ}\text{F}$)	$(^{\circ}\text{C} \times 9/5) + 32$
Meter (m)	Feet (ft)	3.28
Cubic meter (m^3)	Cubic feet (ft^3)	3.531×10^1
Cubic meter (m^3)	Barrel (oil) (bbl)	6.290
Cubic meter (m^3)	Gallon (U.S. liquid) (gal)	2.643×10^2
Cubic meter/second (m^3/s)	Gallon (U.S. liquid)/min (gpm)	1.585×10^4
Watt (W)	Horsepower (electric) (hp)	1.340×10^{-3}
Meter (m)	Inch (in.)	3.937×10^1
Pascal (Pa)	Pound-force/inch ² (psi)	1.450×10^{-4}
Kilogram (kg)	Pound-mass (lb)	2.205
Joule (J)	Watt-hour (Wh)	2.778×10^{-4}

Standard Conditions

$$68^{\circ}\text{F} = 20^{\circ}\text{C}$$

$$1 \text{ atmosphere} = 101,325 \text{ Pascals}$$

PREFIXES

<u>Prefix</u>	<u>Symbol</u>	<u>Multiplication Factor</u>	<u>Example</u>
T	tera	10^{12}	1 Tg = 1×10^{12} grams
G	giga	10^9	1 Gg = 1×10^9 grams
M	mega	10^6	1 Mg = 1×10^6 grams
k	kilo	10^3	1 km = 1×10^3 meters
m	milli	10^{-3}	1 mV = 1×10^{-3} volt
μ	micro	10^{-6}	1 μg = 1×10^{-6} gram

I. INTRODUCTION

Thermal oxidation is a control technology whereby organic vapors are oxidized at high temperatures in the presence of air. Thermal oxidation units have been constructed to control a wide variety of waste gases, the design of the unit depending on the composition and flow rate of the waste gas. The concentration of the volatile organic compounds (VOC) can be converted to the heat generated by the waste gas (heat content) if the specific components and their heats of combustion are known or can be calculated. The heat content range, together with the waste-gas flow rate, determines the design and auxiliary fuel usage.

Three categories of heat contents are used in this report: low (<50 Btu/scf), moderate (50 to 100 Btu/scf), and high (>100 Btu/scf). For waste gases with low heat contents, auxiliary fuel such as natural gas or fuel oil must be added to maintain the combustion temperatures. Heat contents of approximately 13 and 20 Btu/scf in air correspond to 25 and 40% of the lower explosive limit (LEL). Waste gases with heat contents of 20 to 50 Btu/scf (40 to 100% of the LEL) must be diluted with inert gases or be enriched with auxiliary fuel because they exceed the flammable safety limits imposed by insurance companies. Moderate-heat-content waste gases have sufficient heat content for burning but need auxiliary fuel for flame stability.

When the heat content is higher than ~ 100 Btu/scf, the waste gas possesses enough heat value to support a flame by itself and can be considered for use as a fuel gas or boiler feed gas. When flame temperatures resulting from incineration of this type of waste exceed 2200°F , a considerable amount of excess air must be used to cool the unit to 2200°F . Oxidation equipment such as water-wall boilers and high-temperature specialty oxidizers has been successfully designed and operated for temperatures in excess of 2200°F , but is beyond the scope of this study. In many cases the waste gas creating excessive temperatures has been a candidate for flaring.

Conventional thermal oxidizers range in size from a unit capable of controlling several hundred scfm of waste gas to single or multiple units controlling waste gas in excess of 100,000 scfm. Few single thermal oxidizers exist that are

sized for more than 200,000 scfm of flue gas. For a combustion chamber temperature of 1400°F and a waste gas with a heat content of less than 50 Btu/scf and no oxygen, a thermal oxidizer sized for 195,000 scfm of flue gas would handle a waste-gas feed rate of 100,000 scfm. To provide a 1/2-sec residence time for the actual flow rate of 737,000 cfm would require a combustion chamber volume of 6140 ft³. With the length-to-diameter ratio assumed to be 2, this volume would require a refractory-lined cylinder that has an internal diameter of at least 16 ft and that is 32 ft long. Because of shipping size restrictions, larger single units would require field fabrication, which would make the cost much higher. A limit of 100,000 scfm of waste gas was used for this study. Vendors have shop fabricated units up to that size as single units and as multiple units.^{1-4*}

Waste gases containing sulfur or halogens require flue gas scrubbing after thermal oxidation to remove the noxious gases that were formed during oxidation. The scrubbing equipment requires additional capital investment. This analysis is included in the Control Device Evaluation, Thermal Oxidation Supplement (VOC containing halogens or sulfur).

Halogens in the waste gas require high-temperature oxidation to convert the combustion product to a form that can most easily be recovered by scrubbing. For instance chloride-containing waste gases are burned at high temperature to convert the chloride to HCl instead of to Cl₂, since HCl is the more easily scrubbed. The analysis of thermal oxidation of halogenated and sulfonated VOC is also included in the above mentioned report.

*See Sect. VII for all references cited in this report.

II. THERMAL OXIDIZER DESIGN CONSIDERATIONS

A variety of considerations affect the design and selection of the various components comprising a thermal oxidizer system for the control of waste gases: the combustion chamber residence time and temperature, which, in turn, affect the destruction efficiency of the VOC; the auxiliary heat required for flame stability; and the method to be used for heat recovery.

A. RESIDENCE TIME AND TEMPERATURE

Probably the most important considerations in the design of thermal oxidizers are the combustion chamber temperature and residence time. These design variables usually have an impact on both the destruction efficiency and the capital and operating costs of thermal oxidizers.

The combustion temperatures of waste gases vary with the waste-gas heat content. Waste gases with low heat contents will normally have combustion temperatures of 1200 to 1600°F. The thermal oxidizer designer has the option of controlling the combustion temperature by specifying additional auxiliary fuel. Waste gases with moderate heat contents use less auxiliary fuel to support combustion, and the combustion chamber will normally operate in the range of 1600 to 2200°F.

Waste gases with high heat contents determine their own combustion temperatures. The combustion temperatures of these gases can exceed 2200°F, and they are usually satisfactory for use as fuel gases. Figure II-1 shows the feed configurations of thermal oxidizers burning waste gases with low, medium, and high heat contents.

The residence time in the combustion chamber is a design variable specified by the system designer. The combustion chamber is a chemical reactor, and the residence time is the time available for the reaction (oxidation) to occur. Residence times as low as 0.3 sec to several seconds have been utilized in thermal oxidizer designs. Different vendors have defined residence times in different ways. Some include all the available volume of the combustion chamber. Others, however, consider only the volume in which the flue gas is at the combustion temperature, an approach that results in the combustion chamber being larger than it would be if the entire internal volume were used in the calculation of residence time. The fraction of the total volume that is at the combustion

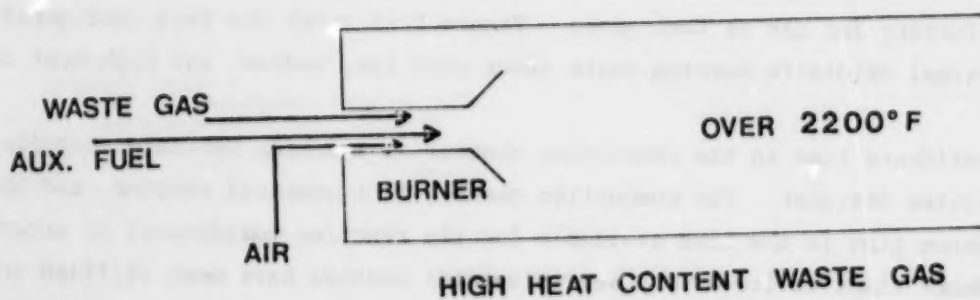
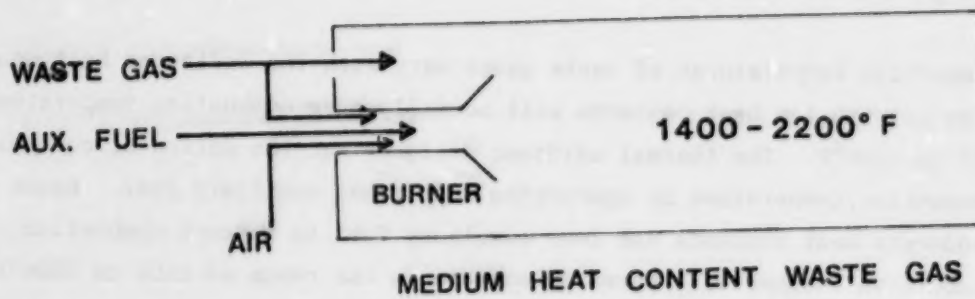
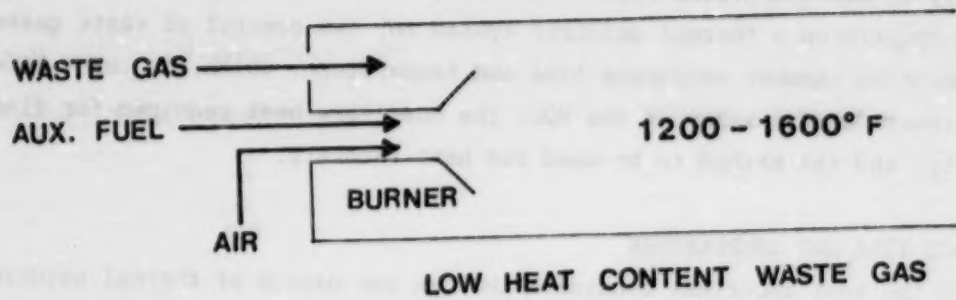


Fig. II-1. Feed Configurations for Thermal Oxidizers Burning Low-, Medium-, and High-Heat-Content Waste Gases

temperature depends on the configuration and design of the flame burner. A burner generating a short flame implies that the waste gas reaches its combustion temperature very rapidly and that nearly the entire internal volume is available for oxidation. A burner developing a long flame has considerably less volume at the combustion temperature than the total internal volume.

Although a design engineer should be quite concerned with the actual burner design and residence time at the combustion temperature, in this control-device evaluation study it is assumed that the entire combustion chamber volume is at the combustion temperature. This is justified since later in this report it is shown that differences as high as 50% in residence time do not significantly affect the annual cost, including capital charges, of thermal oxidation control.

B. VOC DESTRUCTION EFFICIENCY

1. Achievable Destruction Efficiency

The temperatures and residence times of combustion for thermal oxidizers have historically been determined by thermal oxidizer designers using rules of thumb. It is often assumed that thermal oxidizers achieve "complete organic destruction." When specific feed streams required specific destruction data, the chemical manufacturers, their consultants, or the equipment manufacturers would operate pilot units to test burn the real waste in order to determine actual destruction efficiency. This, however, was a costly procedure and was avoided unless there were unusual circumstances. Therefore, except for special cases, VOC destruction efficiency has rarely been measured.

Some vendors have claimed that combustion temperatures several hundred degrees higher than the compound's autoignition temperature should be employed. To date the correlation to destruction efficiencies by this rule of thumb has not been established.⁵

Several other factors are important in determining the destruction efficiency. Residence time at temperature, axial and longitudinal temperature profile, gas density changes in the combustion chamber, and specific component reaction rate (as a function of temperature) all must be known to determine the destruction efficiency, the most important parameter. This requirement implies that for

the thermal oxidizer the combustion chemistry of each component and the physical design of the unit must be known in order to determine the VOC destruction efficiency. Multicomponent oxidation equilibria further complicate the problem.

Recently, some studies have been performed to evaluate this problem. Barnes et al.⁶ have begun a study of the specific kinetics of organic oxidations, but it is theoretical in nature. Followup experimental studies were planned but were cancelled. Lee et al.⁷ have presented a simplified experimental approach to investigate the oxidation kinetics and have given case studies of the destruction relationships of four compounds. In another publication these data are expanded to develop a predictive relationship of VOC destruction. A review of these and other technical resources in this complex area has led to the conclusion that further experimental data are required to ensure that the time, temperature, VOC efficiency relationship is established, which will aid future thermal oxidizer designers. This review has been summarized as a research recommendation for thermal oxidizers and has been submitted to the EPA.⁹

Table II-1 presents the results from tests on five different operating thermal oxidizers and one boiler adapted to burn a waste gas. Some of these tests were run by EPA and some were run by the individual companies who have submitted their data to EPA.^{10,11}

Despite the complexity of the VOC efficiency issue and the need for further work, some estimate of the time, temperature, VOC efficiency relationship is required to design and evaluate thermal oxidizers for this study. Given this need, IT Enviroscience has developed the thermal oxidizer design criteria shown in Table II-2. These criteria are based on engineering experience with a number of VOC applications and the data in Table II-1. In addition, these criteria are based on designing the thermal oxidizer for the specific waste gas under consideration and may involve test burns and pilot unit work with the actual waste stream. Based on these assumptions, it was concluded that a properly operated and individually designed thermal oxidizer would achieve, as a minimum, the stated destruction efficiencies for waste gases with VOC concentration greater than 400 ppmv.

Some of the field unit test data in Table II-1 do not meet the criteria in Table II-2. Analysis of the data lead to the conclusion that insufficient

Table II-1. Results from Actual Thermal Oxidizer Tests

Company	Residence Time/Inlet Flow (SCFM)	Temperature (F°)	Number of Test Runs	Inlet VOC (ppmv) ^a	Outlet VOC (ppmv)	VOC ^b Destruction Efficiency (%)
Union Carbide	2 to 3 sec/ 20,600	1160	6 ^c	11,900	243	96.1
		1475	3 ^c	11,900	10	99.9
Rohm & Haas	1 sec/tank farm vent, (TFV)-12,500 Oxidizer vent, (OXV)- 40,000	1425	3 ^c	TFV 2580 OXV 11,600	1330	82.6
		1510	4 ^c	TFV 2600 OXV 12,800	150	98.3
		1545	1 ^c	TFV 2410 OXV 12,200	25	99.7
Denka	0.6 sec/33,000	1400	3 ^c	950	13	98.5
Monsanto	(Unit size)- 18' dia. X 36' (outlet flow) 75,000	Confidential	Unit 1-6 ^d	Confidential	25	>99
			Unit 2-8 ^d	Confidential	47	>99
Koppers ^e	0.6 sec/30,000	1800	Inlet-4 ^c	850	(Set 1) 7	99.0
			Outlet-6 ^c		(Set 2) 11	97.2
Petro-tex ^f	0.6 sec/14,400	1400	N/A	(Set 1) 10,300 (Set 2) 10,650 (Set 3) 10,300	1000 215 10	70.3 94.1 99.6

^aVOC does not include methane or ethane.

^bVOC destruction efficiency is weight percent.

^cSampling conducted with integrated bags.

^dSampling conducted with grab sample bombs or syringes.

^eThe data in Set 1 and 2 for Koppers were taken during different time periods.

^fInlet and outlet VOC for Petro-tex reported as ppmv methane. This case involves the use of an existing boiler to control a process fume. The data in set one were taken prior to adjustment of the boiler; the data in sets two and three, after adjustment. The specific alterations made by Petro-tex involve changes in mixing induced by retrofit baffles.

mixing or other design deficiency was the most likely cause. In addition, individualized designing with test burns and pilot units adds to costs. In some cases, the costs of this design procedure may be judged too high and the lower efficiencies of units designed for more general application may be acceptable.

As shown by Table II-2, a longer residence time is required to complete combustion when the thermal oxidizer feed contains more than 200 ppm carbon monoxide. This is related to the difficulty encountered in the combustion of carbon monoxide and the tendency for it to disrupt the VOC conversion.

The relationships in Table II-2 show the effect of increased combustion temperatures at similar residence times. They do not show the effect of longer residence times at lower temperatures. As can be seen in Table II-1, excellent destruction efficiencies can be obtained at lower temperatures and longer residence times. Halogenated compounds may be destroyed by using combustion temperatures on the order of those shown in Table II-2 at a residence time of 1 sec or at much elevated temperatures (approaching 3000°F) and shorter residence times. Chemical equilibria between the halogenated compounds, the free halogen, the halogen acid, and the oxygen are functions of the combustion temperatures.³ Since costs are strongly influenced by operating at temperatures in excess of 2200°F, halogenated hydrocarbons need to be handled separately and are discussed in a separate control device evaluation.

The criterion in Table II-2 is based on the assumption that the combustion air fed to the combustion chamber is sufficient to produce 3 mole % O₂ in the flue gas.

The criteria just described are not the only ones for achieving VOC destruction at high levels. For instance, it may be possible to burn waste gases longer (longer residence times) at lower temperatures to achieve the same destruction efficiency. Also, for some easily oxidized compounds, oxidation at, say, 1400°F and 0.5-sec retention time could develop destruction efficiencies above 99%.

Even with the attendant variations this approach is justified for two reasons. First, this criterion is comparable to that used in design and operation of many existing thermal oxidizers and generally represents a conservative design.

Table II-2. Combustion Temperature, Residence Time, and VOC Destruction Relationships

Combustion Temperature (°F)	Residence Time (sec)	VOC Destruction Efficiency ^a (%)
<u>Waste Gas with <2000 ppm Carbon Monoxide</u>		
1400	0.5	>90
1500	0.5	>98
1600	0.5	>99
<u>Waste Gas with >2000 ppm Carbon Monoxide</u>		
1400	0.75	>90
1500	0.75	>98
1600	0.75	>99

^aWith waste gas feeds >400 ppm VOC.

^bThe design of thermal oxidizers to achieve these efficiencies may require test burns and pilot-plant work. In some cases, the costs for this design work may be judged too high and the lower efficiencies of thermal oxidizers designed for less stringent criteria may be acceptable.

Second, the variations of this criterion (that is, different destruction efficiencies at the stated temperatures and residence times) have a small effect on cost effectiveness (cost/lb of VOC destroyed). The major component of the annual cost and energy impact will be related to the combustion temperature and the auxiliary fuel used to achieve that temperature. If a unit is designed to operate at, say, 1600°F and can be shown to achieve a greater VOC destruction than that required by a standard, whether 99% or otherwise, the combustion temperature and the auxiliary fuel usage can be reduced. This conservative design philosophy is similar to the design of most new facilities in the synthetic organic chemicals industry, and the cost impact of any regulations arising from this control device evaluation will almost certainly be conservative. The actual annual cost, cost effectiveness, or energy effectiveness experienced by companies using thermal oxidation for VOC control should be lower than those shown in this report.

2. Calculation of Destruction Efficiency

A secondary problem in the determination of VOC destruction efficiency concerns the method of calculation. VOC destruction efficiency is often calculated by the so-called volumetric efficiency equation:

$$\text{Volumetric efficiency} = 1 - \frac{\text{ppm}_v \text{ in the flue gas}}{\text{ppm}_v \text{ in the waste-gas feed}}$$

which is close to the mass-based destruction efficiency only if the average molecular weight of the VOC in the waste gas is equal to the average molecular weight in the flue gas and if the waste-gas-feed flow is the same as the flue gas flow.

Since these assumptions exist in only a few cases, volumetric efficiencies can be misleading. Mass-based efficiencies are much more desirable. There are, however, some problems in describing thermal oxidation efficiencies in terms of mass flow rates. A thermal oxidizer for VOC destruction receives contaminated VOC gases (sometimes with liquids and solids) and supplementary fuels (natural gas, fuel oils, or high-heat-content waste organic) and burns them with enough air for specified levels of excess or unused oxygen to be achieved in the flue gas. The unburned VOC from both the fuels and the waste gas contribute to the VOC in the flue gas.

In order to establish an efficiency equation to account for the VOC from both sources of flue gas (i.e., the waste gas and the fuel), it would be tempting to define the efficiency as follows:

Total VOC destruction efficiency =

$$1 - \frac{\text{lb of VOC in the flue gas}}{\text{lb of VOC in the waste gas} + \text{lb of VOC in the fuel}}$$

This approach is attractive in that it permits any VOC contribution of the unburned fuel to be assessed directly. It is unacceptable, however, because it does not allow for the evaluation of the degree of destruction of the waste gas. In other words, with VOC destruction efficiencies based on the above equation, it would not be recognized that the thermal oxidizer might burn fuel extremely well but not destroy the waste gas.

When the waste-gas heat content is low, the fuel requirements are high (say, 10 lb of fuel to 1 lb of waste gas VOC). In the above equation the VOC in the flue gas could come from either unburned fuel or waste gas. If the total efficiency required was 90% and the fuel could be burned at 99% (which is easily obtained), then the waste gas could pass through the oxidizer unchanged and the oxidizer would be in compliance with the 90% efficiency requirement. Obviously this approach thwarts the underlying reasons for VOC regulation. A second definition of VOC destruction efficiency therefore must be utilized that bases the efficiency calculation on the waste gas alone:

$$\eta_t(\text{total VOC destruction efficiency}) = 1 - \frac{\text{VOC}_{\text{flue}} (\text{lb of VOC in the flue gas})}{\text{VOC}_w (\text{lb of VOC in the waste gas})}$$

The VOC in the flue gas is comprised of the unburned VOC from the waste gas and the unburned VOC from the fuel, each of which may be expressed in terms of the VOC fed to the thermal oxidizer and destruction efficiency, where the efficiencies of destruction of fuel and VOC in the waste gas are η_{fuel} and η_w , respectively:

$$\text{VOC}_{\text{flue}} = \text{VOC}_w (1 - \eta_w) + \text{VOC}_{\text{fuel}} (1 - \eta_{\text{fuel}})$$

Therefore the VOC destruction efficiency in terms of the waste gas is

$$\eta_t = 1 - \frac{\text{VOC}_w(1 - \eta_w) + \text{VOC}_{\text{fuel}}(1 - \eta_{\text{fuel}})}{\text{VOC}_w}$$

The relationship between η_w , η_{fuel} , and η_t is shown in Fig. II-2 for the case where a waste gas has 21 lb of VOC per hour and 560 lb of fuel per hour is required. This relates to a heat content of about 2 Btu/scf. Also shown on Fig. II-2 is the burner efficiency (η_{fuel}) based on the EPA emission factor for burning natural gas or fuel oil in power plants, industrial boilers, and commercial systems.¹²

The conclusions from Fig. II-1 are that if, say, 99% total efficiency is required and if the EPA fuel oil boiler efficiency is used for the burner efficiency (99.983%), then about 99.4% VOC destruction of the waste gas must be achieved.

Similarly, if for any reason the burner efficiency drops from 99.983% to 99.960%, then the 99% total efficiency could not be achieved by using the above definition of efficiency. This is particularly significant in view of the qualifications in the EPA emission factor table¹² indicating that surges, upsets, turndown, or poor design or maintainance could increase the emission factors significantly. From these statements it could be concluded that the 99.960% fuel efficiency level would be difficult to maintain on a continuous basis. Overall efficiency levels approaching 99% based on the waste gas demand the best fuel burner designs and the smoothest possible operation.

A final consideration is that, by present definition, methane (natural gas) is not considered to be a VOC. The VOC destruction efficiency based on the waste gas is independent of burner efficiency if the fuel is natural gas. It has become apparent that for a variety of reasons natural gas may not be the dependable choice for supplementary fuel.

C. FLAME STABILITY

An auxiliary-fuel minimum of 5 Btu/scf of waste gas for medium-heat-content waste gases is assumed in this study. For medium- to high-heat-content waste gases, auxiliary fuel amounting to 10% of the waste-gas heat content is added for flame stability. A waste gas with a heat content of 100 Btu/scf thus requires 10 Btu/scf of auxiliary fuel. Obviously, for very high heat content waste gases

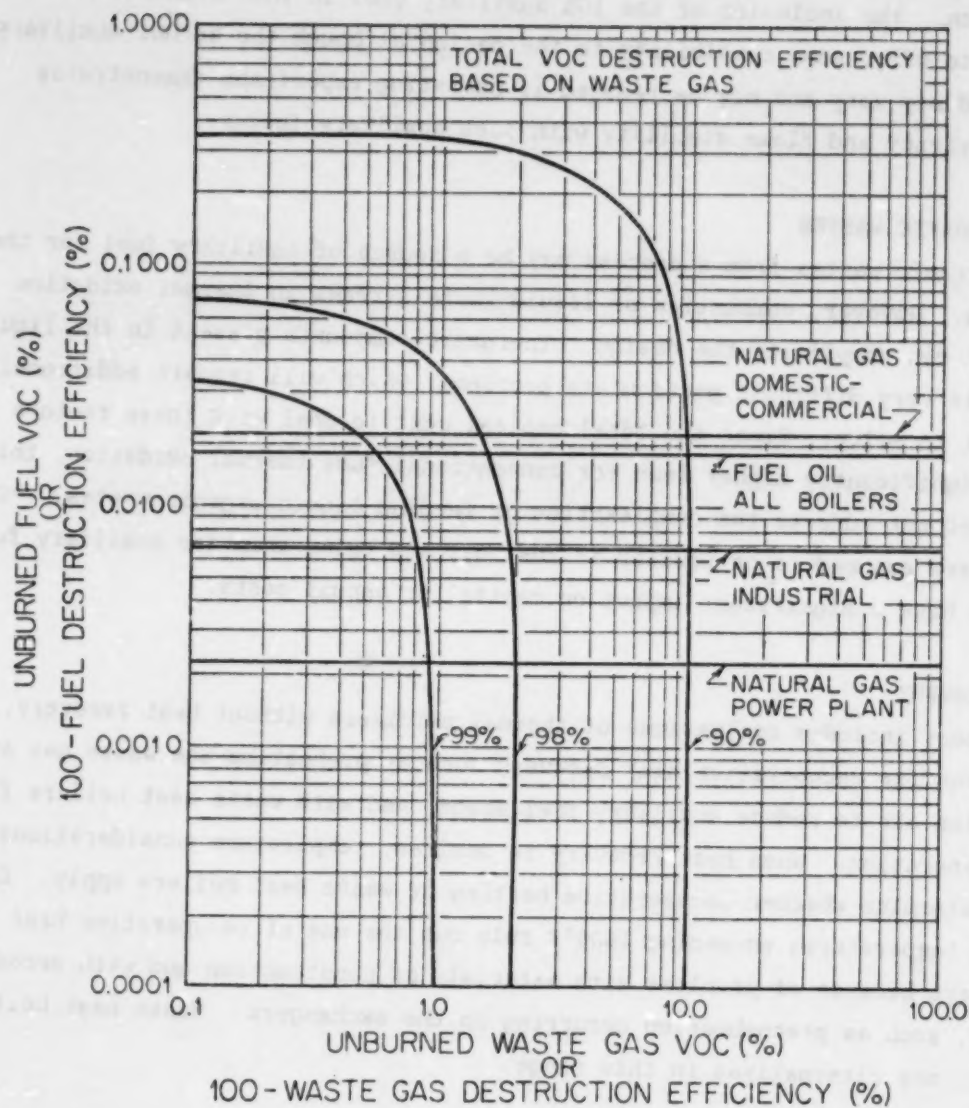


Fig. II-2. Relationship Between Total VOC Destruction Efficiency Based on Waste Gas, Fuel Destruction Efficiency, and Waste Gas Destruction Efficiency for a 2-Btu/scf-Heat-Content Waste Gas

(>400 Btu/scf) the auxiliary fuel requirements for flame stability diminish. The actual selection of auxiliary fuel for specific thermal oxidizer designs is highly judgmental and requires consideration of several factors specific to the application. The inclusion of the 10% auxiliary fuel in this evaluation is intended to offer some credibility to the design although the actual auxiliary fuel added may vary and may be reduced if operating experience demonstrates good efficiency and flame stability with less auxiliary fuel.

D. LIQUID ORGANIC WASTES

Liquid organic wastes from a process may be a source of auxiliary fuel for thermal oxidation. However, combustion of liquid waste streams in thermal oxidation equipment can complicate the design. Inorganic compounds present in the liquids can create very difficult particulate problems, which will require additional equipment to solve. Since the total capital cost to deal with these factors can be significantly higher than for conventional fume thermal oxidation, this study does not address the complexities of feeding liquid organic wastes. "Clean" liquids are assumed to be similar to fuel oil and using them for auxiliary fuel will not have a significant impact on capital or annual costs.

E. HEAT RECOVERY

This report includes evaluations of thermal oxidizers without heat recovery, with means for recuperative heat recovery such as preheating the waste gas and combustion air to reduce auxiliary fuel usage, and with waste heat boilers for steam generation. When heat recovery is desired, temperature considerations could determine whether recuperative heating or waste heat boilers apply. Combustion temperatures exceeding 1600°F rule out the use of recuperative heat exchangers because of problems with materials of construction and with secondary factors, such as precombustion occurring in the exchangers. Waste heat boilers, however, are alternatives in this range.

Recuperative heating is possible with temperatures between 1500 and 1600°F only if the part of the exchanger closest to the flame is manufactured from special materials (nickel alloys, etc). Temperatures less than 1500°F are compatible with standard recuperative heater designs. Waste heat boilers may be considered throughout all ranges. For waste gases with moderate or high heat contents, recuperative heating is not an option since it is of value only when used to reduce the auxiliary-fuel requirement. Waste heat boilers may not be installed at those production locations where additional process steam cannot be used.

III. BASIS FOR THERMAL OXIDIZER DESIGN

After the design considerations described in the preceding section were evaluated, those sensitive design variables that would have an effect on cost and energy were identified and are described below.

A. EFFECT OF SENSITIVE DESIGN VARIABLES ON COST AND ENERGY

A distinction must be made between those design variables which, if changed by a small amount, would cause significant changes in capital annual or energy costs. These are called sensitive variables, and the cost curves given later in the report generally include them as parameters. Other variables may be quite important for an individual system design but have minor effects on economic or energy impact conclusions.

The approach used in this study was to determine the sensitivity of certain variables by means of computerized heat and material balance calculations. Through this process, estimates of the relationships between the variables and equipment design and operating costs may be derived. Primary variables that are a function of the waste gas are the waste gas temperature, pressure, flow rate, VOC composition and VOC average molecular weight, VOC carbon, oxygen, and hydrogen (and other component) ratios, VOC heats of combustion, and the nitrogen (and other inert gases), oxygen, and water contents, and the presence of special contaminants (particulates, halogens, high levels of sulfur).

The waste gas temperature is assumed to be 100°F for the base case, but an increase or decrease within reasonable boundaries will have little effect upon the capital or operating costs and it is therefore not a significant variable. Sensible heat carried by the waste gas is small compared to that required to raise its temperature to the combustion conditions. About 3.5 Btu/scf is required to increase the temperature of nitrogen from 80°F to 260°F. This compares with fuel heat requirements on the order of 60 to 80 Btu/scf to raise the waste gas to combustion temperatures. Waste gas temperature differences within this range could not change the total heat requirements by more than about 6%. Waste gas pressure is assumed to be 1.5 psig. Pressure changes within reasonable limits of 1.5 psig also will have no significant effect on capital or operating costs.

Flow rate is a very significant variable for both capital and operating costs. The waste gas flows shown in the figures throughout this report are translated to scfm of waste gas to the thermal oxidizer.

Heat content of the waste gas is a significant variable. VOC molar concentration; average molecular weight; carbon, hydrogen, and oxygen ratios; and heats of combustion (Btu/lb of VOC) are all expressed in the variable of the heat content of the waste gas (Btu/scf). By assessing the heat of combustion of the VOC being destroyed and the mole % VOC concentration in the waste gas, the heat content can be determined (Btu/scf) as shown by the family of compound lines on Fig. III-1. Multicomponent VOC systems may be described when the mole fractions of each component are known. The contribution of carbon monoxide to the total heat content may also be analyzed in this way. Table III-1 gives VOC molar heats of combustion.¹³

If the actual flue gas composition is needed, a component material balance must be performed, for which carbon, hydrogen, and oxygen ratios are required. In order to estimate "typical" values for those ratios, 219 organic compounds containing C, H, O, N, and Cl were surveyed.¹⁴ Table III-2 summarizes this information based on groupings of different classes. The VOC component averages of 68.3% carbon, 11.4% hydrogen, and 20.3% oxygen were used to establish heat value plus heat and material balance for this evaluation. Compounds containing chlorine or sulfur are not included in the impact assessment of this study but covered in the Control Device Evaluation, Thermal Oxidation Supplement (VOC containing halogens or sulfur).

The level of oxygen in the waste gas is important because compounds containing high amounts of oxygen lessen the level of combustion air required and reduce the total thermal oxidizer flue gas, which affect both capital and operating costs. No additional combustion air is required when the oxygen in the flue gas exceeds 3 mole %. This leads to a smaller unit and lower capital and operating costs. Although some designers will use various levels of oxygen to determine combustion requirements, the value of 3 mole % is based on accepted practice and will be constant for all the design calculations in this evaluation. In order to generate a conservative design and impact analysis, the waste gas in this report is assumed to have no oxygen. Maximum combustion air is therefore

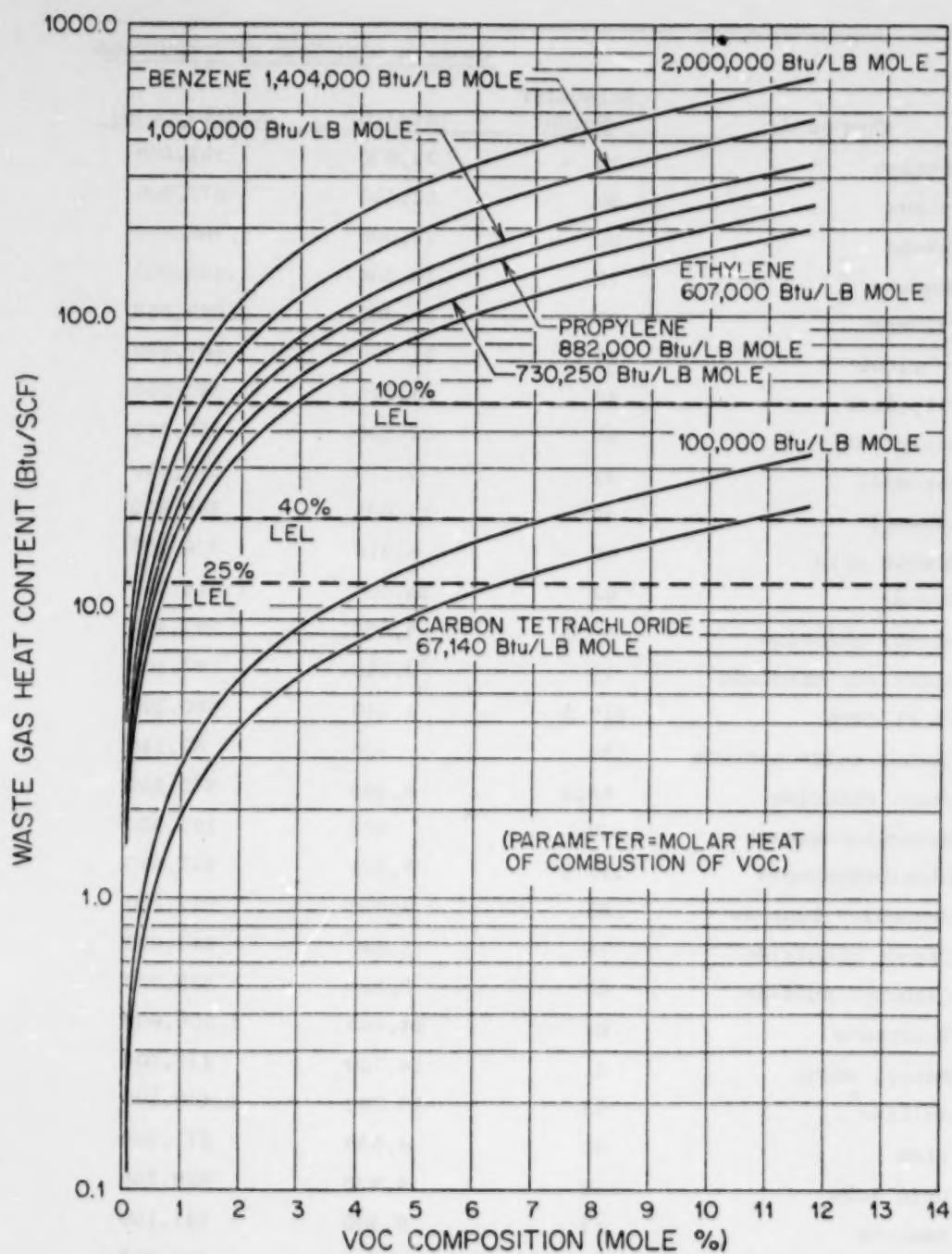


Fig. III-1. Relationship Between Waste Gas Heat Content, VOC Composition, and VOC Molar Heat of Combustion

Table III-1. VOC Molar Heats of Combustion*

Compound	Molecular Weight	Gross or High-Heat of Combustion	
		(Btu/lb)	(Btu/lb-mole)
Methane	16	23,900	382,400
Ethane	30	22,400	672,000
Hexane	86	20,800	1,788,800
Benzene	78	18,000	1,404,000
Toluene	92	18,300	1,683,600
Ethylene	28	21,700	607,600
Propylene	42	21,000	882,000
Acetylene	26	21,500	559,000
Methanol	32	9,770	312,600
Ethanol	46	12,800	588,800
Acetic acid	60	6,270	376,200
Phenol	94	14,000	1,316,000
Methyl chloride	50.5	5,850	295,400
Methylene chloride	85	2,260	192,100
Chloroform	119.5	1,340	160,100
Carbon tetrachloride	154	436	67,140
Ethyl chloride	64.5	8,840	570,200
Hexachloroethane	237	835	197,900
Dichlorobenzene	111.5	8,230	917,600
Hexachlorobenzene	285	3,220	917,700
Carbon disulfide	76	5,840	443,800
Carbonyl sulfide	60	3,920	235,200
Thiophene	84	14,400	1,209,600
Methyl amine	31	14,700	455,700
Aniline	93	15,700	1,460,100
Urea	60	4,530	271,800
Uric acid	168	4,930	828,200
Ammonia	17	8,300	141,100
Nitromethane	61	5,000	305,000
Nitrobenzene	123	10,800	1,328,400
Trinitrobenzene	213	5,610	1,194,900

*See ref 12.

Table III-2. Summary of Organic Compound Components Surveyed

Class		Amount (wt %)				
		C	H	O	N	Cl
Carbon, hydrogen compounds only	Low value*	79.9	7.7			
	Mean	85.8	14.2			
	High value*	92.3	20.1			
Carbon, hydrogen compounds and carbon, hydrogen, oxygen compounds	Low value	27.0	2.1	0		
	Mean	68.3	11.4	20.3		
	High value	92.3	20.4	7		
Carbon, hydrogen, and chlorine compounds only	Low value	7.8	0			31.5
	Mean	34.3	4.7			60.6
	High value	64.0	9.8			92.2
Carbon, hydrogen, oxygen, and nitrogen compounds and carbon, hydrogen, and nitrogen compounds	Low value	26.1	3.7	0	13.0	
	Mean	60.1	11.3	3.6	25.0	
	High value	78.5	15.4	52.0	60.8	
All compounds	Low value	7.8	0	0	0	0
	Mean	62.5	10.4	17.3	3.9	5.9
	High value	92.3	20.1	20.1	60.8	92.2

*Low and high values correspond to the compound with the lowest and highest percentage of each element (e.g., carbon). Different compounds apply to different elements; therefore the values shown will not be additive.

assumed. When ample oxygen in a low or medium heat content waste gas is available, the size of the combustion chamber to control the waste gas may be as low as one-half the volume as shown in this report.

At 1600°F and 14.7 psia the heat capacity of water vapor is about 11.8 Btu/(lb-mole)(°F). The heat capacity of air under the same conditions is 7.7 Btu/(lb-mole)(°F).¹⁵ Since saturated conditions are assumed for the waste gas in calculations in this report, the water content in the flue gas is at a maximum unless entrained liquid water droplets enter with the feed. Auxiliary-fuel requirements can increase significantly in this case since the heat capacity of the flue gas increase and the heat of vaporization for water, 18,000 Btu/lb-mole of water, must be supplied. However, this can normally be avoided with proper design.

The presence of special contaminants can have significant effects on capital and operating costs. A different design criterion is required with halogenated feeds. Other contaminants, such as sulfur, could require the addition of absorbers to the control device. Investigation of these special cases are assessed in the Control Device Evaluation, Thermal Oxidation Supplement.

A second group of significant variables relates to the design criterion of thermal oxidation. Combustion chamber temperature and residence time must be specified as functions of the feed components and VOC destruction efficiency. Combustion chamber temperature has a significant effect on fuel costs for feeds with low heat contents. Feeds with moderate heat contents determine their own combustion temperature, whereas feeds of high heat content will markedly increase the size of the equipment if the combustion chamber must be controlled at levels much below their normal combustion temperatures. The addition of excess air or water to reduce the combustion temperature results in a larger flue gas flow and larger equipment to maintain the same residence time. Residence time has a major effect on capital costs but a relatively small effect on operating costs. Even though a larger unit has a larger heat loss by radiation, it is normally on the order of a few percent and does not materially increase the operating cost.

The significant variables investigated in this report are waste gas flow, heat content (encompassing a variety of composition-related variables), combustion

temperature, residence times, and destruction efficiencies. Evaluation of these parameters using generally worst-case or conservative assumptions will lead to economic and energy impact conclusions that will equal or exceed actual operating costs and energy impacts for the applicable cases. This approach, then, leads to conservative economic and energy costs on which to base future regulations.

B. PROCEDURE USED FOR DESIGNING THERMAL OXIDIZER SYSTEM

The design procedure that was used for the thermal oxidizer unit of this study was developed based on the above variables.

1. Combustion Chamber

The volume and composition of the waste gas were first determined. The heat content of the gas was determined by the identification of the VOC components (including CH_4 and CO) and the use of Table III-1 (plus other molar heats of combustion data) and Fig. III-1. The minimum combustion temperature and residence times were taken from Table II-2, depending on the presence of carbon monoxide. Supplementary fuel is required for low-heat-content waste gas to maintain the desired combustion chamber temperature and flame stability. The supplementary fuel requirement as a function of waste-gas heat content is shown by Fig. III-2. The waste gas is assumed to have no oxygen and therefore the combustion air required is at a maximum. With these data the total combustion flue gas (in scfm) from all sources, including the waste gas, auxiliary fuel, combustion air, and combustion products, was calculated for the waste-gas heat content range (Btu/scf) displayed in Fig. III-2. To correlate with the following design, size, and cost projections for the full range of waste gas rates, the flue gas flow of Fig. III-4 is expressed as a ratio of flue gas to waste gas as a function of waste-gas heat content. In waste gases with high levels of oxygen the ratio of flue gas to waste gas approaches 1. The more conservative case of a waste gas with no oxygen as shown in Fig. III-3 is used in this evaluation.

Conversion of scfm to actual cubic feet per minute (acfm) is necessary for sizing of the combustion chamber. Figure III-4 shows this relationship (based on the ideal gas law). Standard conditions assumed throughout this report are 32°F and 760 mm Hg and their equivalents. The ratio of acfm to scfm is read from Fig. III-4 and is multiplied by the combustion chamber flue gas in scfm. Until

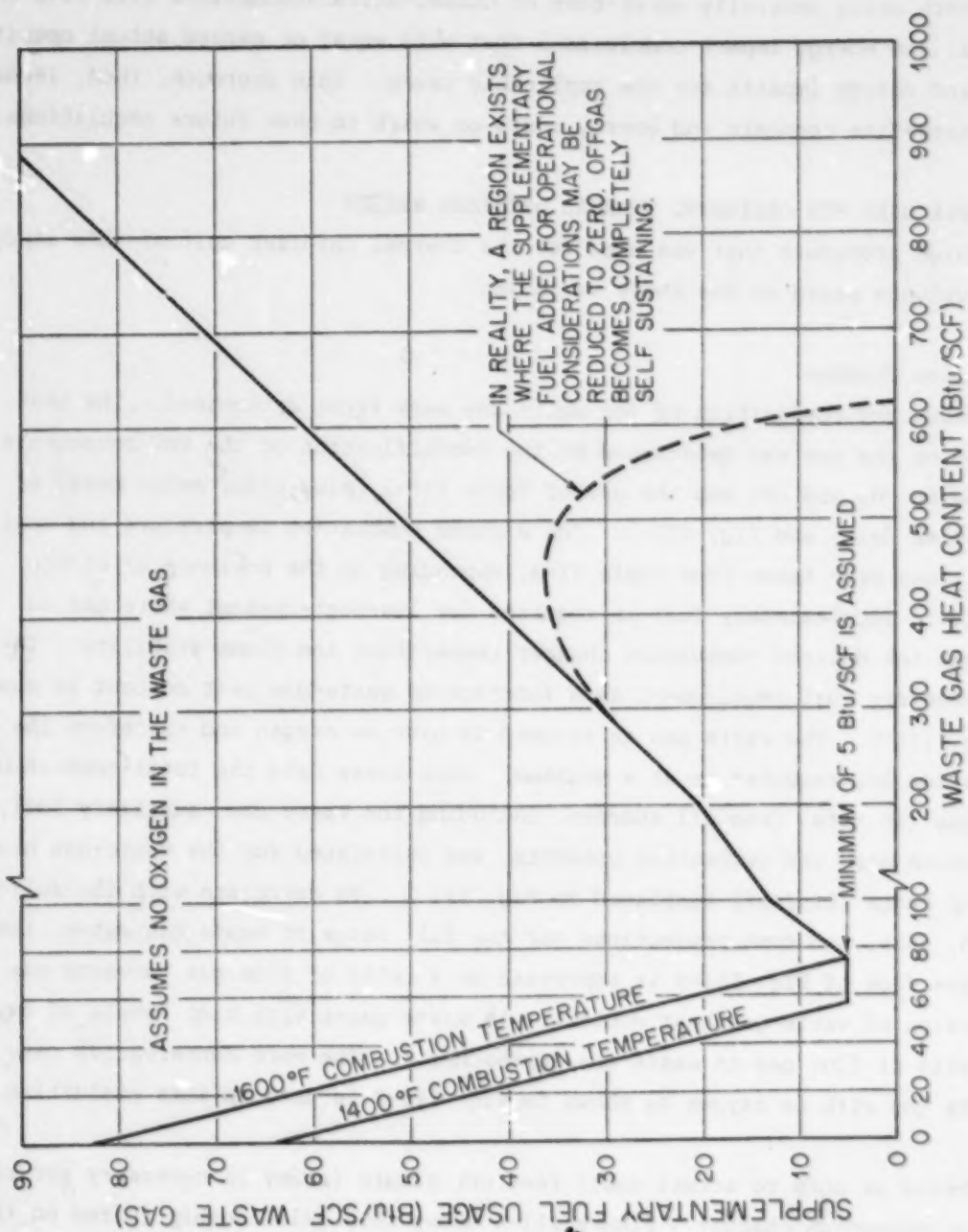


Fig. III-2. Supplementary Fuel Usage vs Waste Gas Heat Content

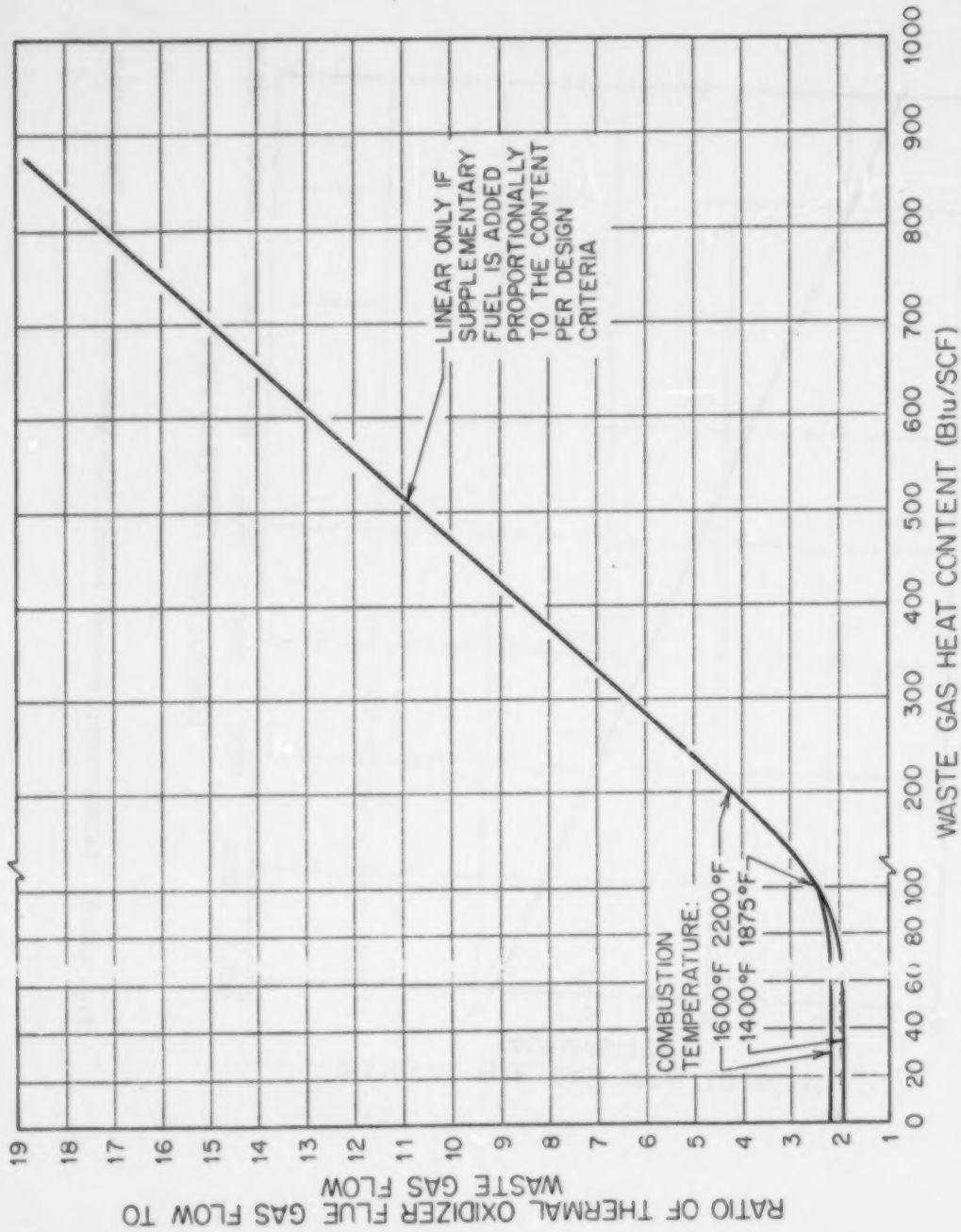


Fig. III-3. Ratio of Thermal Oxidizer Flue Gas Flow to Waste Gas Flow vs Waste Gas Heat Content

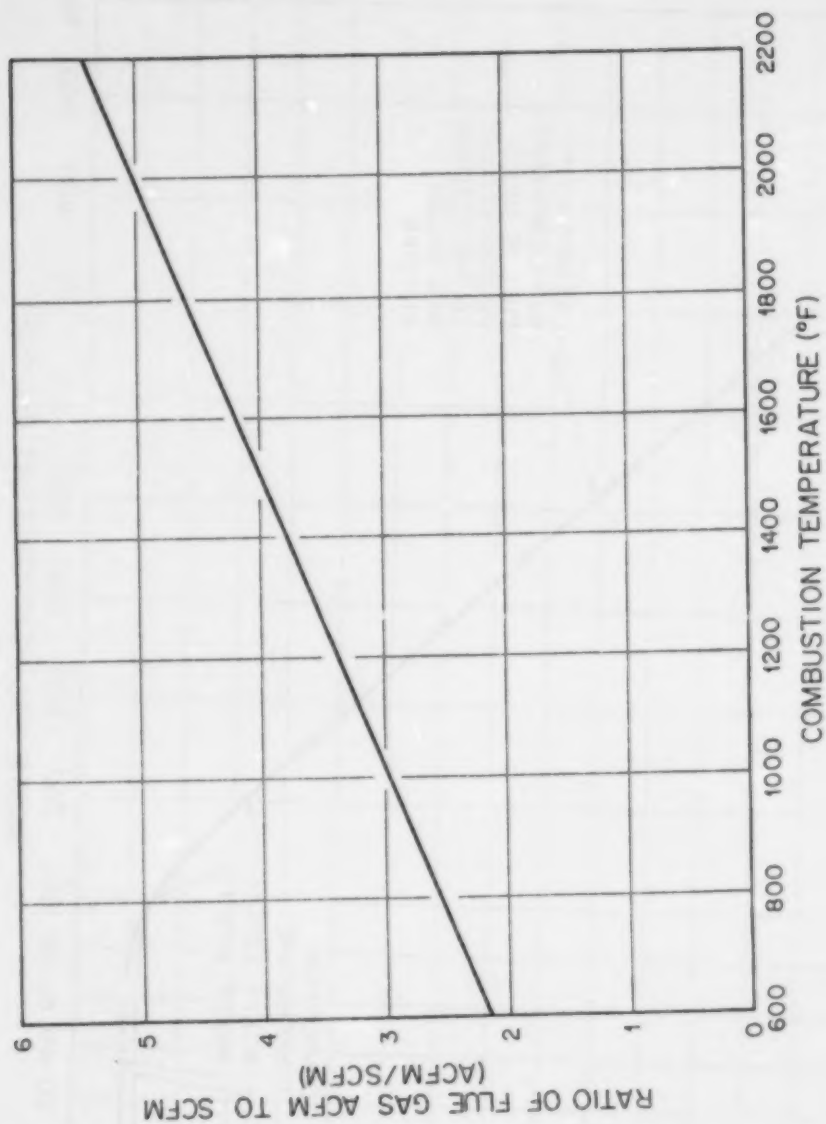


Fig. III-4. Relationship Between Actual Flow Rates and Flue Gas Temperature

this point in the report, the term heat content has referred to the energy generated by combustion of compounds in the waste gas. Heat content when used in relation to the flue gas in this report refers to the energy contained by the hot gases because of their temperature. The basis used for the heat recovery calculations is the heat content of the flue gas as a function of temperature, as shown in Fig. III-5. The dotted lines in Fig. III-5 correspond to a reasonable variation in specific heats or heat capacities of the flue gas.

Residence times (sec) from Table II-2 are used to calculate the combustion chamber internal volume. The combustion chamber flue gas flow (in acfm) is converted to actual cubic feet per second and then multiplied by the residence time (in sec) to determine the combustion chamber internal volume in cubic feet.

2. Fans

Fans for the waste gas and combustion air are both specified. The flow rates of the waste gas and combustion air and the pressure drops of the thermal oxidizers are used to calculate fan sizes. Pressure drops of 6 in. of water were assumed for the thermal oxidizer with 0 to 30% heat recovery; 8 in. was assumed for thermal oxidizers with 50% heat recovery; and 10 in. was assumed for thermal oxidizers with 70% heat recovery or waste heat boilers. The waste-gas fan capacity is based on the waste-gas flow rate. Table III-3 gives the combustion air flow/waste gas ratio as a function of waste-gas heat content. The relationship is based on desired combustion temperatures, as previously discussed, and is used to size the combustion air fan. The combustion air volume and fan size may be determined by the use of this ratio multiplied by the volume of waste gas. Waste gases containing significant levels of oxygen reduce the combustion air required and reduce the size of the combustion air fan. Waste gases generated at higher pressure often need no fan.

3. Recuperative Heat Recovery

A recuperative heat recovery system transfers heat from the flue gas into the waste gas and combustion air, thus lowering the requirements for auxiliary fuel. However, since less fuel is required for a given volume of waste gas, less combustion air is also needed to burn the fuel and the resulting flue gas volume is less than that of a thermal oxidizer without recuperative heat recovery. Since the flue gas volume is lower, the amount of recoverable heat is also lower.

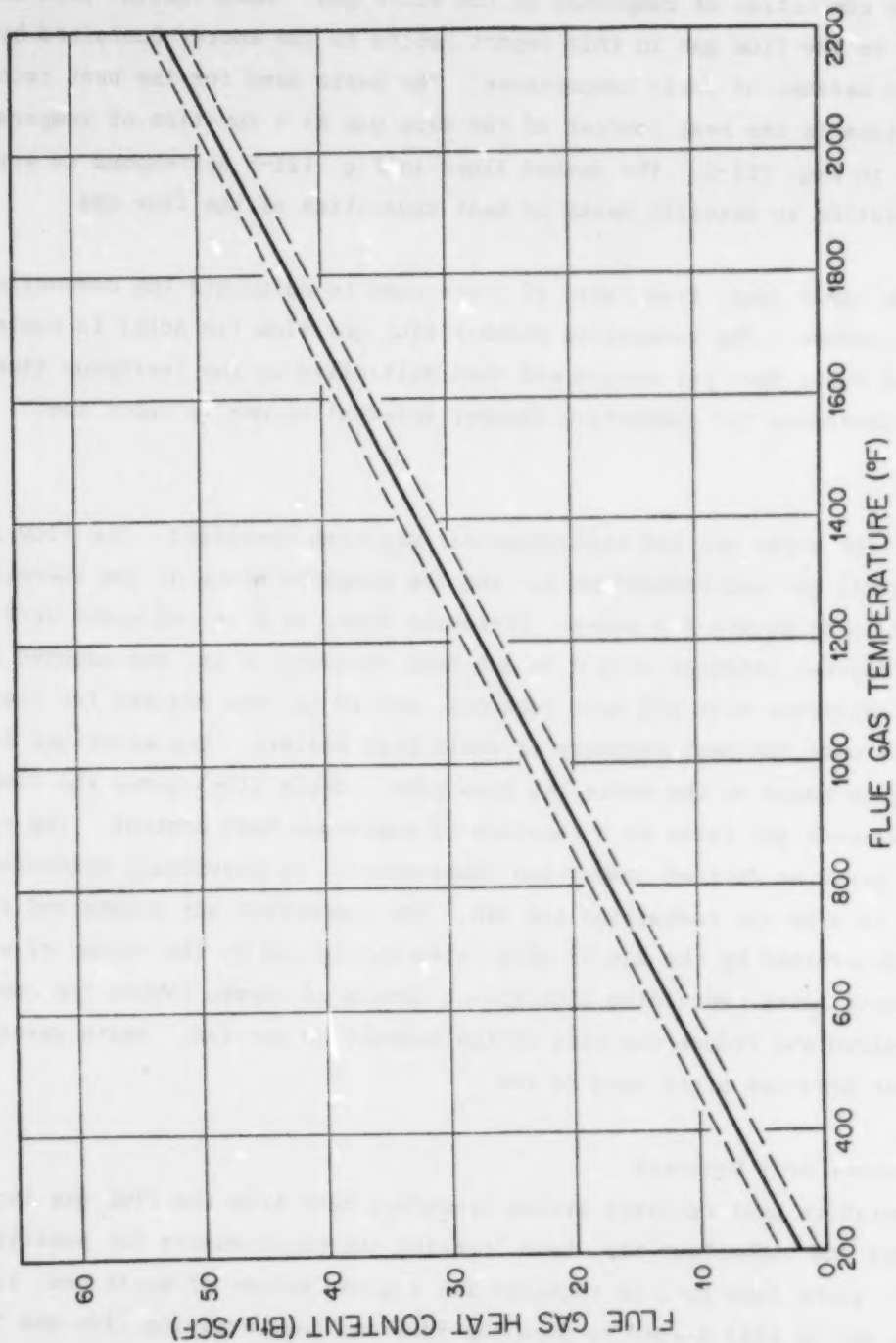


Fig. III-5. Flue Gas Heat Content

Table III-3. Ratio of Combustion Air to Waste
Gas Flow Rate vs Waste Gas Heat Content

Waste Gas Heat Content (Btu/scf)	Combustion Temperature (°F)	Combustion Air to Waste Gas Flow Ratio* ($\frac{\text{scf of combustion air}}{\text{scf of waste gas}}$)
2	1400	0.87
2	1600	1.1
100	1875	1.4
200	2200	3.1
400	2200	7.3

*Thermal oxidizer conditions:

No oxygen in waste gas.

VOC molar heat of combustion = 730,250 Btu/lb-mole.

VOC molecular weight = 50.

VOC C, H, O fraction = 68.3 wt % C, 11.4 wt

% H, 20.3 wt % O.

Average waste gas molecular weight = 29.

Water content of combustion air = 1.0 wt %.

3 mole % O₂ in flue gas after oxidation.

The combustion chamber, combustion air fan, recuperative heat exchanger area, and stack all vary directly with flue gas volume. Recuperative heat recovery has the effect of shrinking the entire system, depending on the level of heat recovery. Table III-4 gives the factors that describe this variation of system size with various levels of recuperative heat recovery. The factors were developed with the computerized heat and material balance for 1400°F and 1600°F combustion temperatures and a waste gas with no oxygen. When the waste gas contains significant levels of oxygen, the size-reduction effect as related to recuperative heating diminishes. Although some size reduction is still seen, the factors in Table III-4 approach 1.

The fuel used per volume of waste gas is reduced when recuperative heat recovery is used compared to that when no heat recovery or waste heat boilers are used. Table III-5 gives the fuel reduction factor for various heat contents, combustion temperatures, and levels of recuperative heat recovery. These factors, multiplied times the fuel usage values from Fig. III-2, give the actual fuel usage when recuperative heat recovery is used. A minimum of 5 Btu/scf of fuel was used for this evaluation.

The heat content of the flue gas from the combustion chamber is essentially a function of temperature. Thermal oxidation flue gases from waste-gas incineration will have similar compositions of N_2 , O_2 , H_2O , and CO_2 and therefore similar specific heats. The level of heat recovery is defined as

$$1 - \frac{\text{heat content of flue gas after heat recovery}}{\text{heat content of flue gas before heat recovery}}$$

With it assumed that the temperature rise in the waste gases plus the combustion air in a recuperative exchanger is equal to the temperature drop of the flue gases in the exchangers, a design relationship may be developed relating the heat exchanger surface area to the level of heat recovery. This assumption is normally a realistic one since the mass flows and heat capacities of the waste gas plus air and those of the flue gas are normally similar. Any amount of heat recovery is possible up to the point where the temperature of the preheated combustion feed gas approaches the flue gas temperature. This point represents the maximum possible heat recovery for a combustion chamber at a given temperature. The heat content of the flue gas is shown in Fig. III-5. Figures III-6

Table III-4. Thermal Oxidizer Size Reduction Factor for
Recuperative Heat Recovery Systems

Level of Recuperative Heat Recovery (%)	Size Reduction Factor	
	1400°F	1600°F
30	0.850	0.830
50	0.770	0.744
70	0.701	0.667

Table III-5. Fuel Reduction Factors for
Recuperative Heat Recovery Systems

Level of Recuperative Heat Recovery (%)	Fuel Reduction for Various Waste Gas Heat Contents			
	1 Btu/scf	10 Btu/scf	13 Btu/scf	20 Btu/scf
<u>Combustion Temperature = 1400°F</u>				
30	0.635	0.525	0.490	0.400
50	0.420	0.260	0.210	0.061
70	0.240	0.044		
<u>Combustion Temperature = 1600°F</u>				
30	0.590	0.530	0.501	0.380
50	0.393	0.297	0.260	0.134
70	0.212	0.083	0.034	

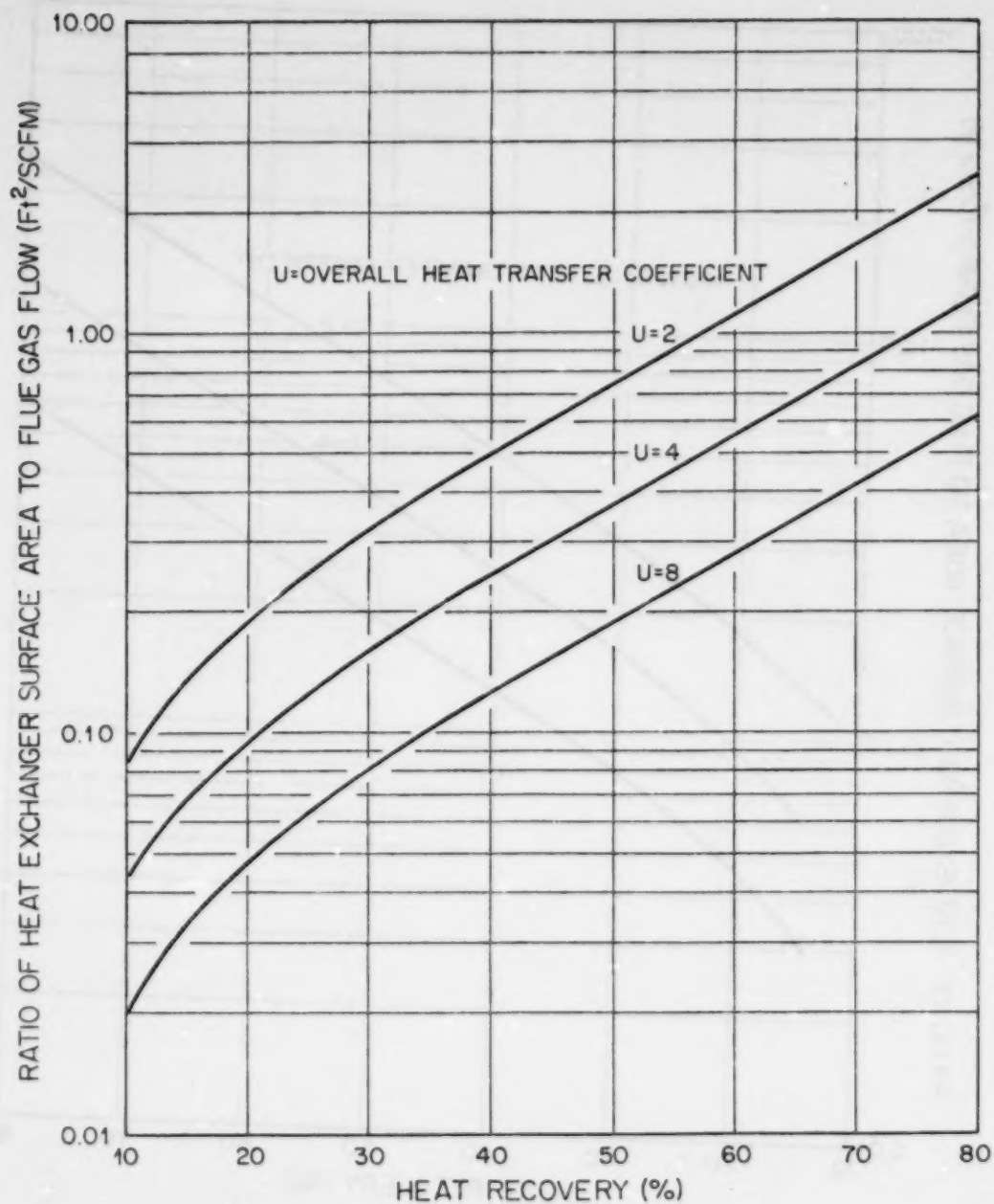


Fig. III-6. Recuperative Heat Exchanger Design at 1400°F Combustion Temperature (100°F Feed Gas, Constant Specific Heat)

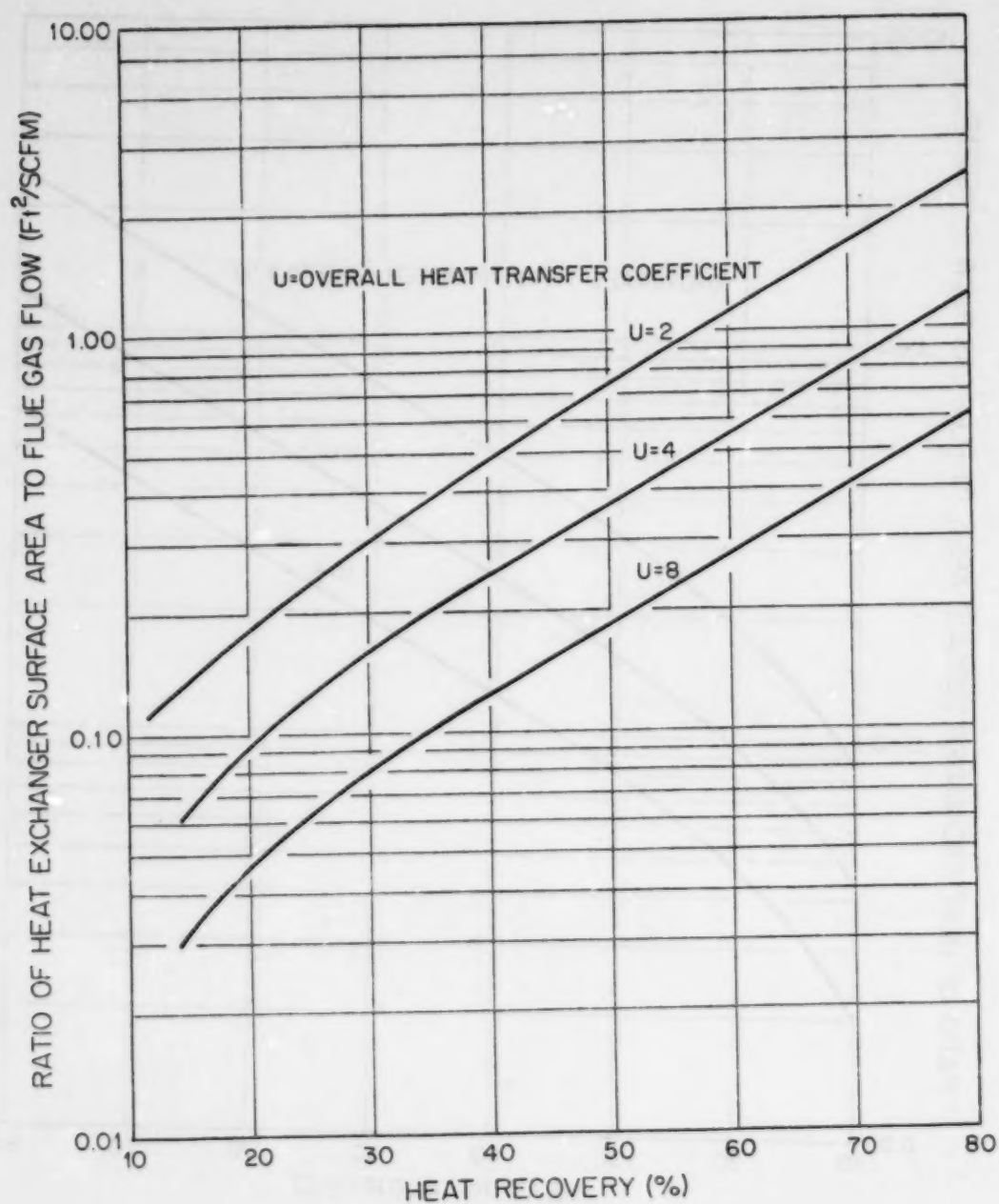


Fig. III-7. Recuperative Heat Exchanger Design at 1600°F
Combustion Temperature (100°F Feed Gas, Constant Specific Heat)

and 7 show the ratio of the heat exchanger surface to the flue-gas flow rate as functions of the percent heat recovery and the overall heat transfer coefficient (U) for a recuperative exchanger for 1400°F and 1600°F flue gases.

Overall heat transfer coefficients of these heat exchangers depend on the type of design. Cross-flow exchangers tend to have overall heat transfer coefficients around 2 to 5,^{2,3} whereas U-tube exchangers are reported to have coefficients of 6 to 8.⁴ Design calculations leading to cost estimates in this report assume an overall heat transfer coefficient of about 4.

Once a value for the area-to-flow ratio is determined, the heat exchanger area is calculated by multiplying the flue gas flow (in scfm) by the ratio.

4. Waste Heat Steam-Generation Boiler Heat Recovery

The maximum heat recovery by a waste heat steam-generation boiler depends on the temperature of the exhaust gas after it exits the boiler. Although flue gas temperatures can be reduced to the condensation temperatures of the components in the flue gas, in this study it is assumed that the flue gas exhaust temperature is 500°F. Since more heat could be recovered, this is a more conservative but more universal assumption. With the flue gas exhaust temperature specified, the maximum heat recovery is established. This relationship is shown in Fig. III-8. The dotted lines in this figure relate to a reasonable range of heat capacities or specific heats of the flue gas. Figure III-9 relates the ratio of boiler-tube surface area to the flue-gas flow rate and the flue gas temperature, steam temperature, and overall heat transfer coefficient. The surface area of the waste heat boiler is determined by multiplying the flue-gas flow rate by the ratio from Fig. III-9. The flue-gas flow rate from a thermal oxidizer employing a waste heat boiler is the same as that from a thermal oxidizer using no heat recovery. Unlike the recuperative heat recovery case, no reduction in system size or fuel use is seen when a waste heat boiler is used.

5. Stack

The cross-sectional area of the stack is determined by assuming a superficial linear velocity (3000 fpm) and dividing into the actual flow rate of the flue gas or exhaust gas (with or without heat recovery). Figure III-4 can be used to convert the scfm to acfm when the temperature is known. Table III-6 lists

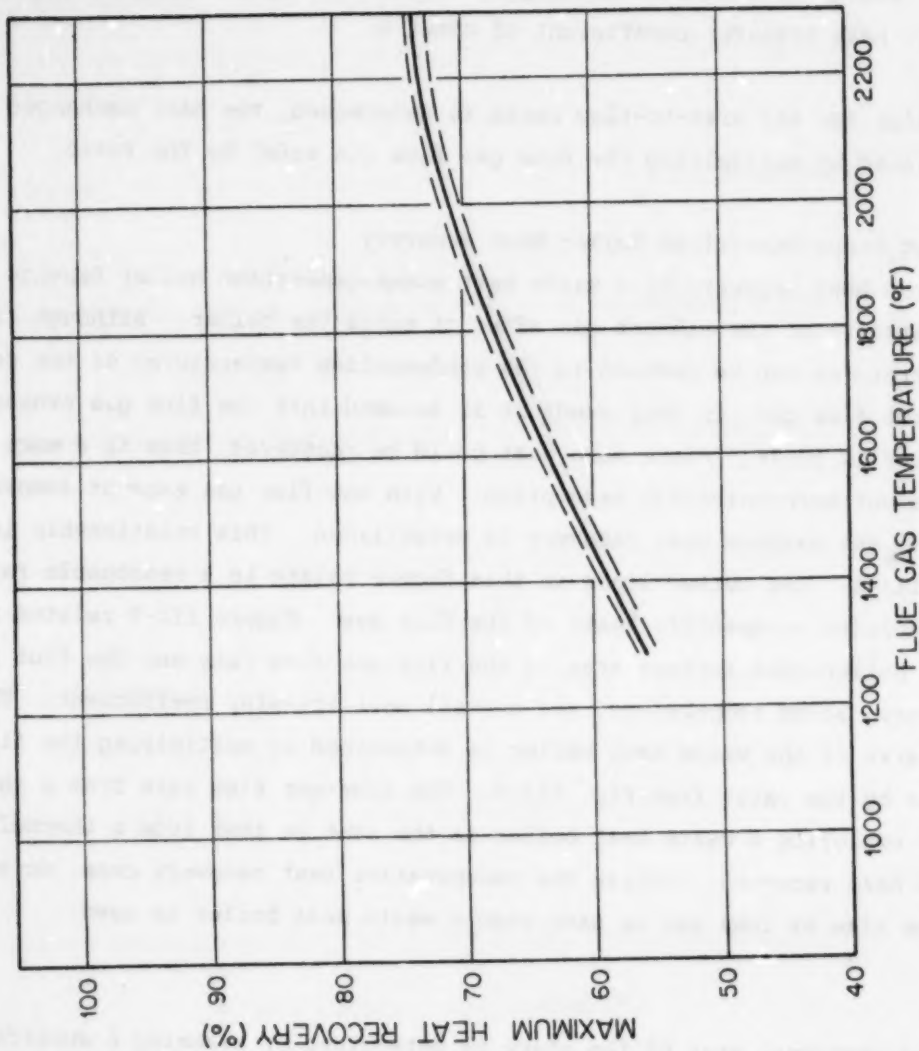


Fig. III-8. Maximum Heat Recovery from a Waste Heat Boiler (Exhaust Temperature = 500°F)

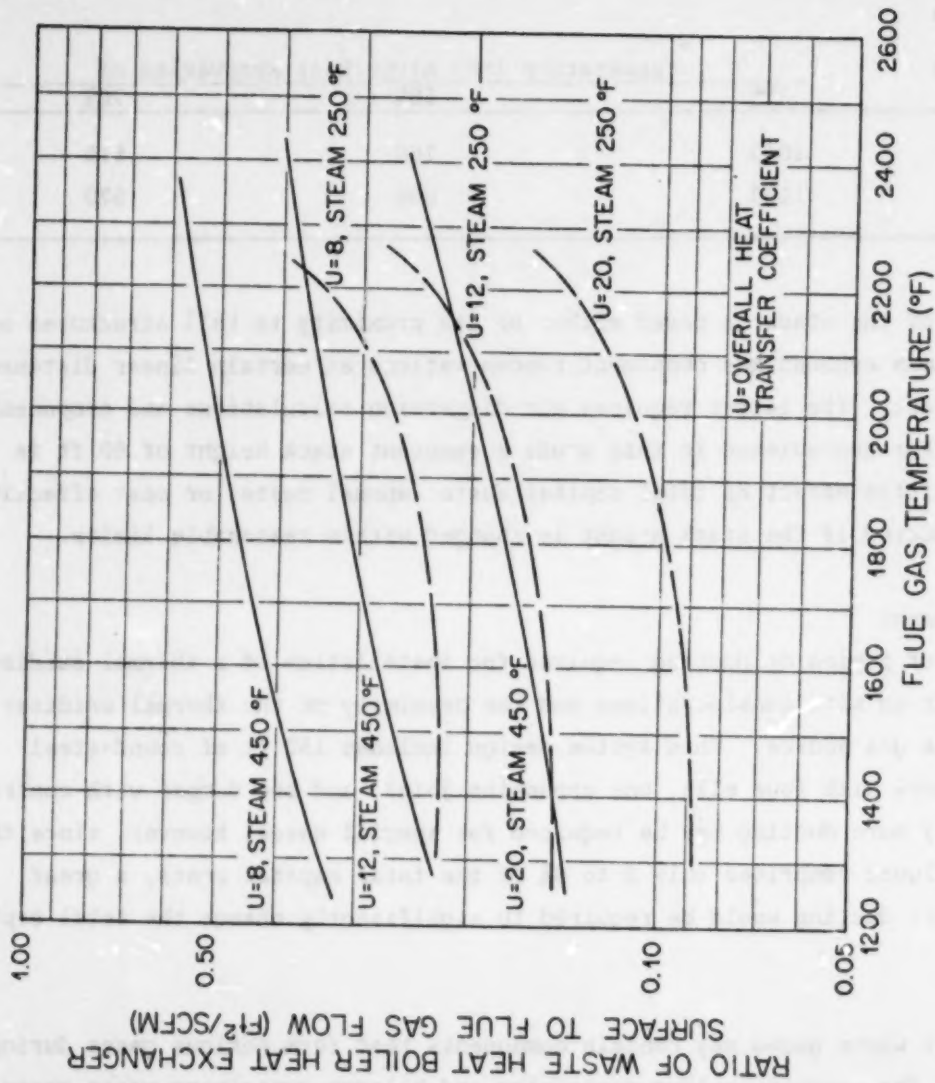


Fig. III-9. Ratio of Waste Heat Boiler Heat Exchange Surface to Flue Gas Flow vs Flue Gas Temperature

the temperatures of the gases after they are exhausted from the recuperative heat exchangers or from the waste heat boiler. A waste heat boiler operating at maximum heat recovery will have an exhaust gas temperature of 500°F.

Table III-6. Flue Gas Exhaust Temperature After Heat Recovery

Combustion Chamber Temperature (°F)	Temperature (°F) After Heat Recoveries of		
	30%	50%	70%
1400	1040	760	510
1600	1160	860	570

The height of the stack is based either on its proximity to tall structures or on the maximum exhaust gas component concentrations at certain linear distances from the stack. The latter requires air dispersion calculations and component criteria. For convenience in this study a constant stack height of 80 ft is assumed. Little effect on total capital costs, annual costs, or cost effectiveness is expected if the stack height is changed within reasonable limits.

6. Other Equipment

The amount of piping or ducting required for installation of a thermal oxidizer is dependent on site considerations and the proximity of the thermal oxidizer to the waste gas source. This system design includes 150 ft of round-steel inlet ductwork with four ells, one expansion joint, and one damper with control. Considerably more ducting may be required for special cases; however, since the ducting included comprises only 2 to 8% of the total capital costs, a great deal of extra ducting would be required to significantly change the total capital cost.

Some special waste gases may contain components that form noxious gases during combustion, for example, sulfur-containing and halogen-containing waste gases. Flue gas scrubbers are required to reduce the level of these noxious gases in the exhaust gas. The Control Device Evaluation, Thermal Oxidation Supplement addresses the incremental capital increase when this option is required.

C. COMBUSTION ALTERNATIVES

An alternative technology that is used for incineration of waste gases with low heat contents is catalytic oxidation, in which a catalyst is used to increase the oxidation rate of the gases at lower temperatures. This is discussed in the Control Device Evaluation, Catalytic Oxidation.

Waste gases with high heat contents are often candidates for use as fuel gas to a boiler or are controlled by flaring. A control device evaluation report for this technology is also in preparation.

IV. CONSIDERATIONS FOR INSTALLING THERMAL OXIDATION CONTROL EQUIPMENT

Thermal oxidizers can be large process units, depending on the volume of waste gas to be controlled, and could require a plot of land as large as 300 ft by 300 ft for installation.

Since thermal oxidizers utilize combustion with a flame for achieving VOC destruction, the unit must be located at a safe distance from process equipment using flammable chemicals or special designs must be employed to minimize the risk of explosion.

Thermal oxidizers require natural gas or fuel oil, electrical power, and instrument air and, if scrubbing is needed, water at the site. If steam is generated from waste heat, then it is useful to minimize the distance from the waste heat boiler to the steam consuming site. No special utilities are needed.

Retrofitting thermal oxidizers into existing plants requires careful consideration of site location since all the above considerations apply and sufficient space in an existing plant may not be available. The unit may have to be located at longer distances from the waste gas source than would be required for a new plant.

V. COST AND ENERGY IMPACTS OF THERMAL OXIDIZERS

A. COST BASIS

The estimated capital costs for total systems combinations and for various components are presented in this section. These estimated costs represent the total investment, including all indirect costs such as engineering and contractors' fees and overheads, required for purchase and installation of all equipment and material to provide a facility as described. These are "battery limit" costs and do not include the provisions for bringing utilities, services, or roads to the site, the backup facilities, the land, the research and development required, or the process piping and instrumentation interconnections that may be required within the process generating the waste gas feed to the thermal oxidizer.

The estimated costs are based on a new plant installation; no retrofit cost considerations are included. Those costs are usually higher than the cost for a new site installation for the same system and include, for example, demolition, crowded construction working conditions, scheduling construction activities with production activities, and longer interconnecting piping. Since these thermal oxidizer systems require a relatively large land area and the safety aspects of an open flame are an important factor, the longer interconnecting piping will probably be the most significant of these retrofit cost factors. These factors are so site specific that no attempt has been made to provide costs. For specific retrofit cases, rough costs can be obtained by using the new site data and adding as required for a defined specific retrofit situation.

The method used to develop these estimated capital costs was based on preliminary vendor quotes for the purchase of major equipment items or from such sources as Richardson Engineering Co. data, and factoring up to installed costs based on the factors presented in Table V-1. The expected accuracy of the total installed cost with this degree of engineering detail using this factor method is $\pm 30\%$.

This method of obtaining estimated total installed capital costs is suitable for a cost study or for screening estimates. Table V-1 lists the factor ranges used for various cost components and is based on historical data of IT Environmental Process Engineering.

Table V-1. Factors Used for Estimating Total Installed Costs

A = Major Equipment Purchase Cost Plus 0.1 to 0.35 Allowance

Installation Costs

Foundations	0.06A + \$100 X number of pumps
Structures	0.15A (no structures) to 0.30A (multideck structures)
Equipment Erection	0.15A to 0.30A (depending on complexity)
Piping	0.40A (package units) to 1.10A (rat's nest)
Insulation	0.06A or 0.15 X piping (normal) to 0.30 X piping (bulk hot or cold)
Paint	0.05A
Fire Protection	0.01A to 0.06A (depending on requirements)
Instruments	0.10A to 0.30A or 0.01A to 0.25A + \$50,000 to \$300,000 for process control computer
Electrical	0.15A or 0.05A + \$500 per motor

B = Base Cost	A + Sum of Installation Costs
Sales Tax	0.025A + 0.025B
Freight	0.16A
Contractor's Fees	0.30 (B-A)

C = Total Contract	B + Taxes, Freight, and Fees
Engineering ^a	0.01C to 0.20C
Contingencies ^b	0.15C

D = Process Unit Installed Cost C + Engineering + Contingencies

E = Total Subestimates Sum of semidetalled subestimates (buildings, site development, cooling towers, etc.). Each subestimate should include taxes, freight, fees, engineering and contingency, and should be escalated to date of expenditure for that cost component. Engineering costs, contingencies, and escalation factors for these subestimates will vary according to the type of job.

F = Total Project Cost D + E

^a Includes cost from capital project teams, process engineering, engineering, purchasing, and other support groups.

^b Contingency should not be applied to any cost component that has been committed by either purchase order or contract.

The estimate is based on the purchase cost of major equipment A, including a 10 to 35% allowance for other equipment and an assessment of the quality of vendor quotes. A 10% allowance is used for project definition that includes process flow sheets and specific budget quotes and a 35% allowance flow sheet for block flow sheet definition and generalized equipment quotes or prices.

B. CAPITAL COSTS

1. Thermal Oxidizer Complete Systems

Figures V-1 through V-6 show the total estimated capital costs that were obtained for various system combinations and operating conditions, such as various waste-gas heat contents, residence times, operating temperatures, etc. Combinations and conditions are as defined on each figure. The method for estimating the total system costs consisted of combining appropriate costs for components established as described below and of adding minimum site development, estimating allowance, and nominal vendor startup costs. The total added costs for site development, estimating allowance, and vendor startup were then prorated back to each individual component. The individual component cost curves can therefore be combined to build up the total cost for any complete system desired. Since the component cost curves include a portion of the costs normally assigned for final complete system installations, they are quite specific for use in estimating thermal oxidizer systems and therefore should not be used indiscriminately as a general cost estimating reference for the individual components. Curves showing purchase costs of combustion chambers, recuperative heat exchangers, and waste heat boilers are presented in Appendix A.

2. Combustion Chamber

Preliminary purchase quotes for thermal oxidizer combustion chambers were obtained from vendors. Six vendors were contacted by telephone and letter and three of them were visited to clarify details and costs: Combustion Engineering Air Preheater, Peabody Engineering, and Hirt Combustion Engineers. The installed costs shown in Fig. V-7 through V-11 were obtained by factoring the preliminary quotes as described above.

The information furnished by vendors included thermal oxidizer purchase cost vs capacity for several of their standard units. The units were quoted as being

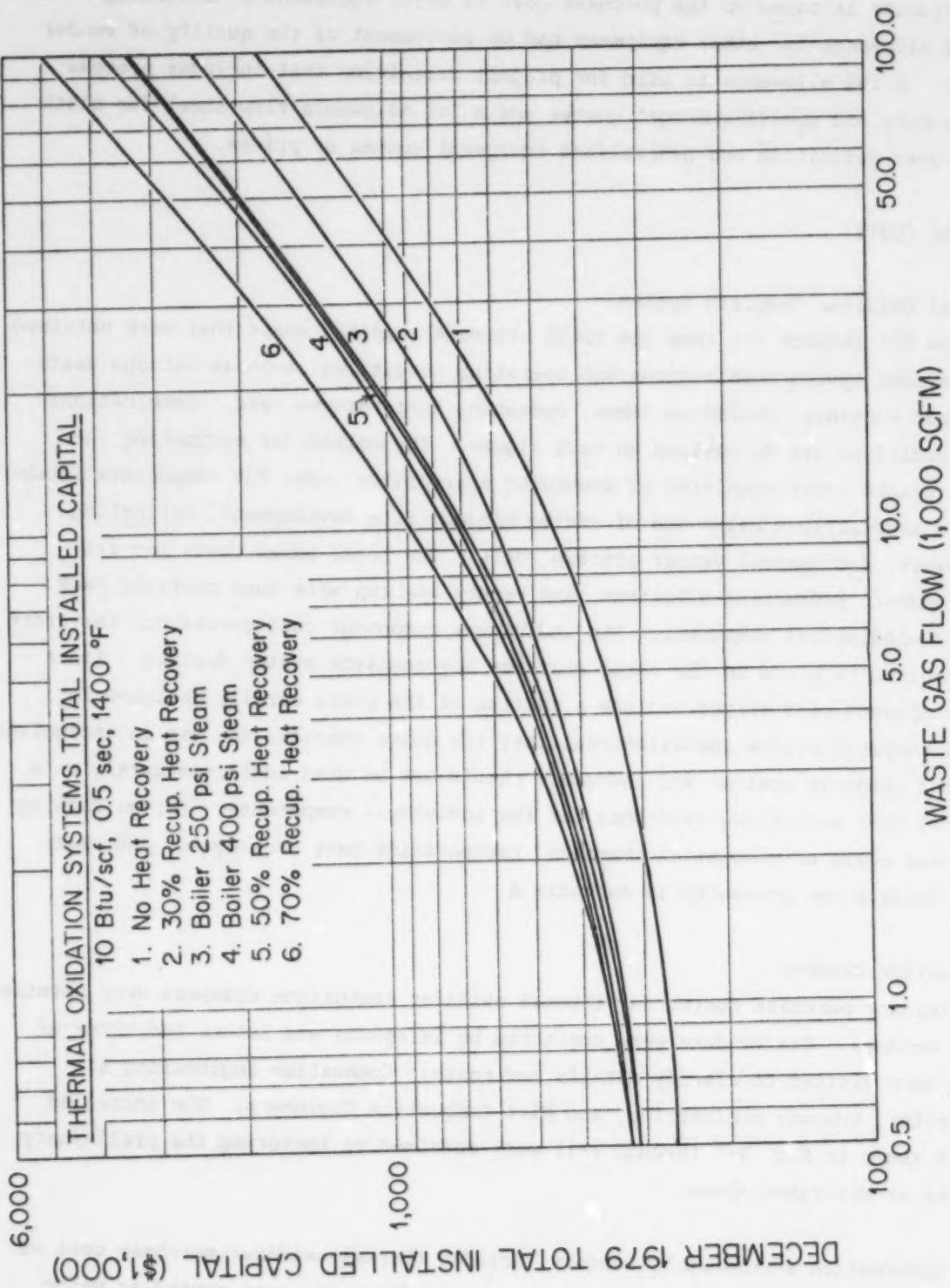


Fig. V-1. Total Installed Capital Cost for Thermal Oxidation Systems with Waste-Gas Heat Content = 10 Btu/scf, Residence Time = 0.5 sec, and Combustion Temperature = 1400°F

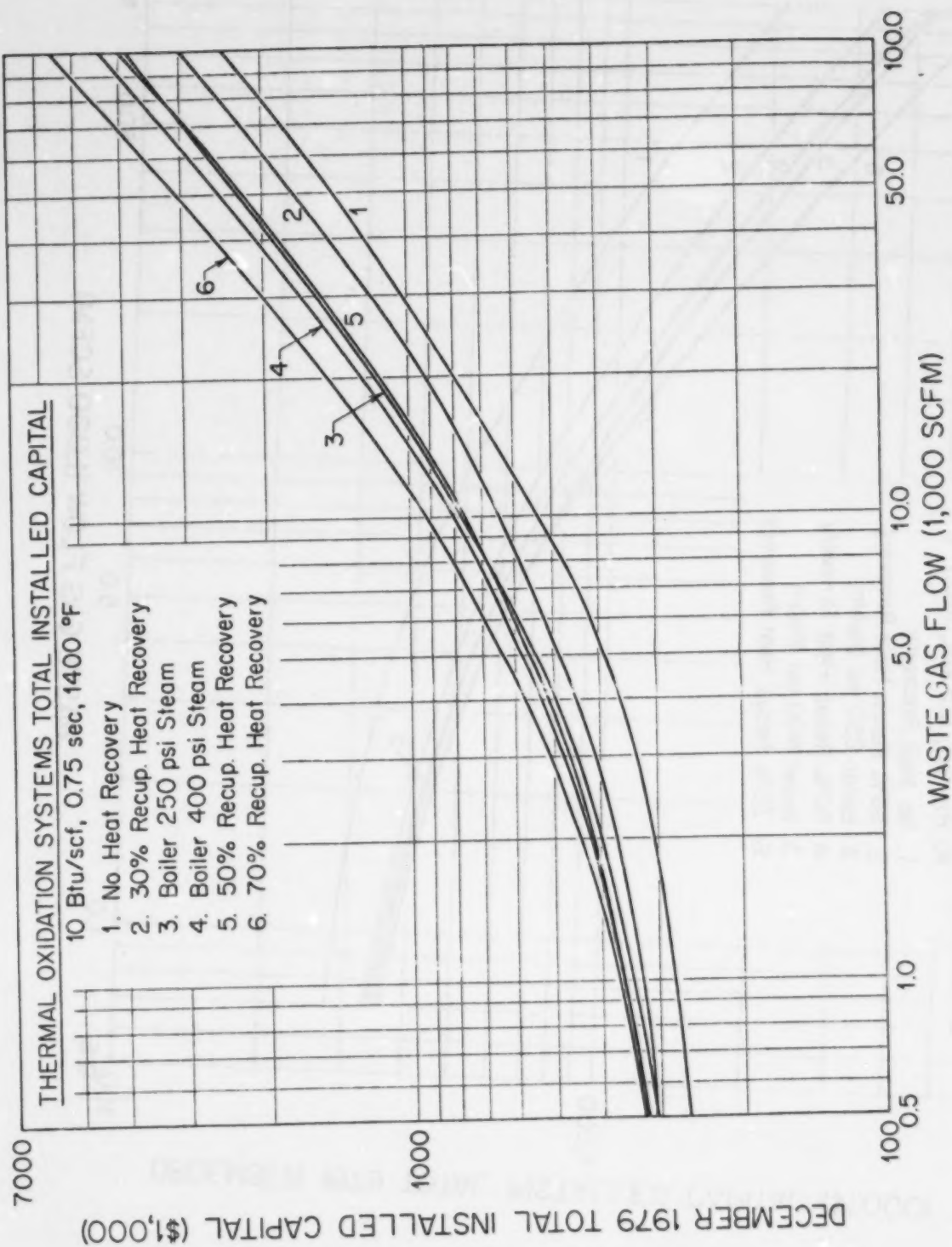


Fig. V-2. Total Installed Capital Cost for Thermal Oxidation Systems with Waste-Gas Heat Content = 10 Btu/scf, Residence Time = 0.75 sec, and Combustion Temperature = 1400°F

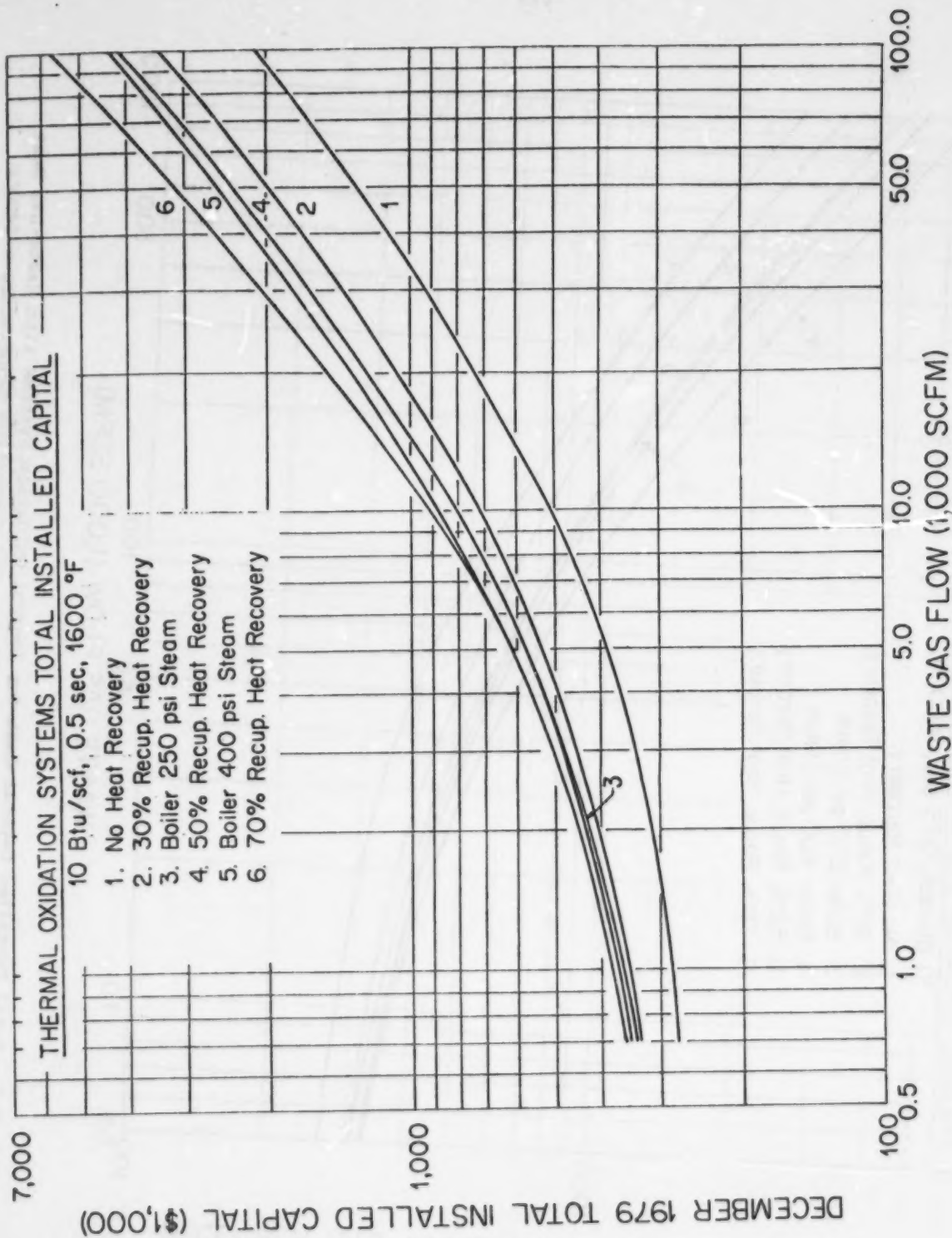


Fig. V-3. Total Installed Capital Cost for Thermal Oxidation Systems with Waste-Gas Heat Content = 10 Btu/scf, Residence Time = 0.5 sec, and Combustion Temperature = 1600°F

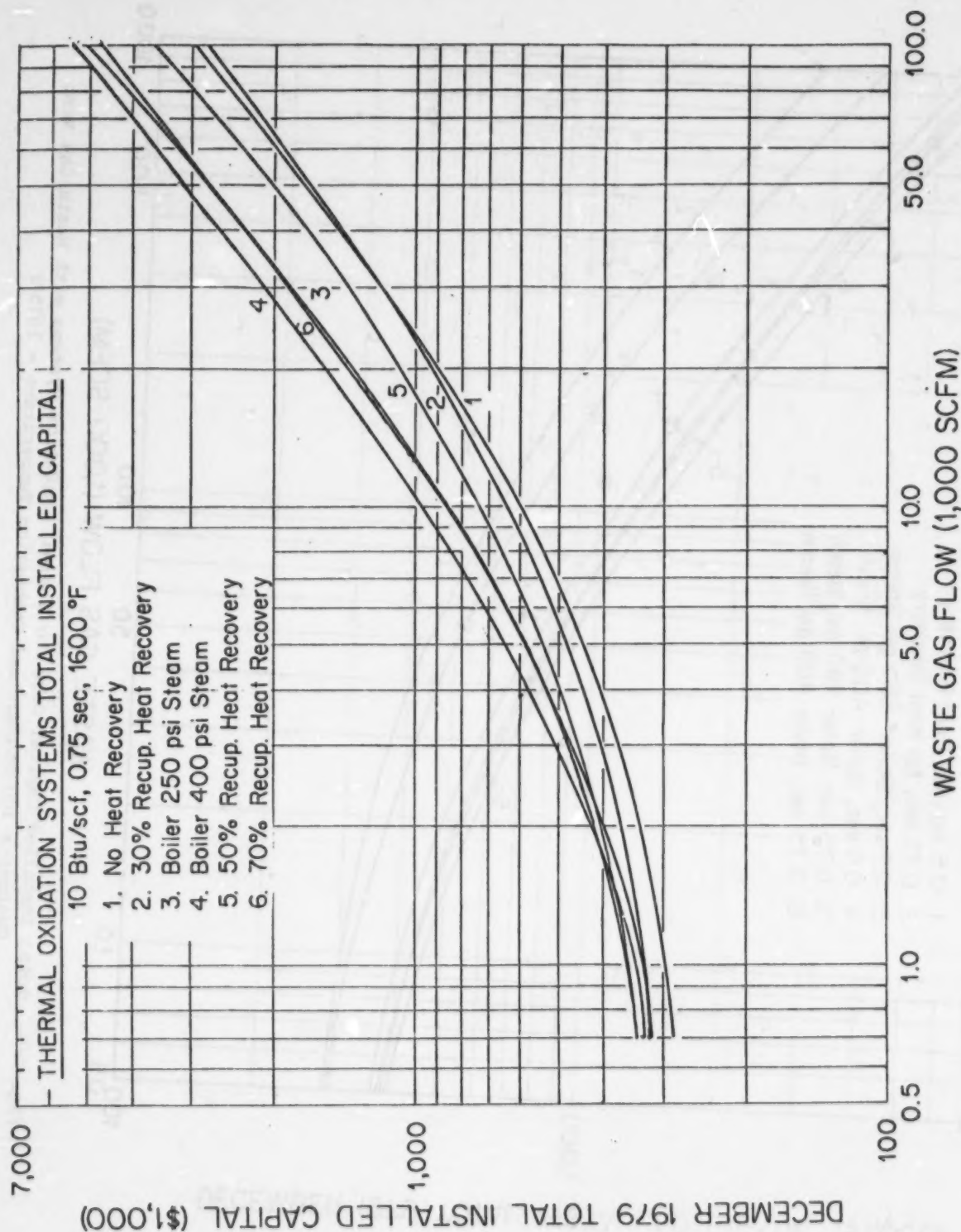


Fig. V-4. Total Installed Capital Cost for Thermal Oxidation Systems with Waste-Gas Heat Content = 10 Btu/scf, Residence Time = 0.75 sec, and Combustion Temperature = 1600°F

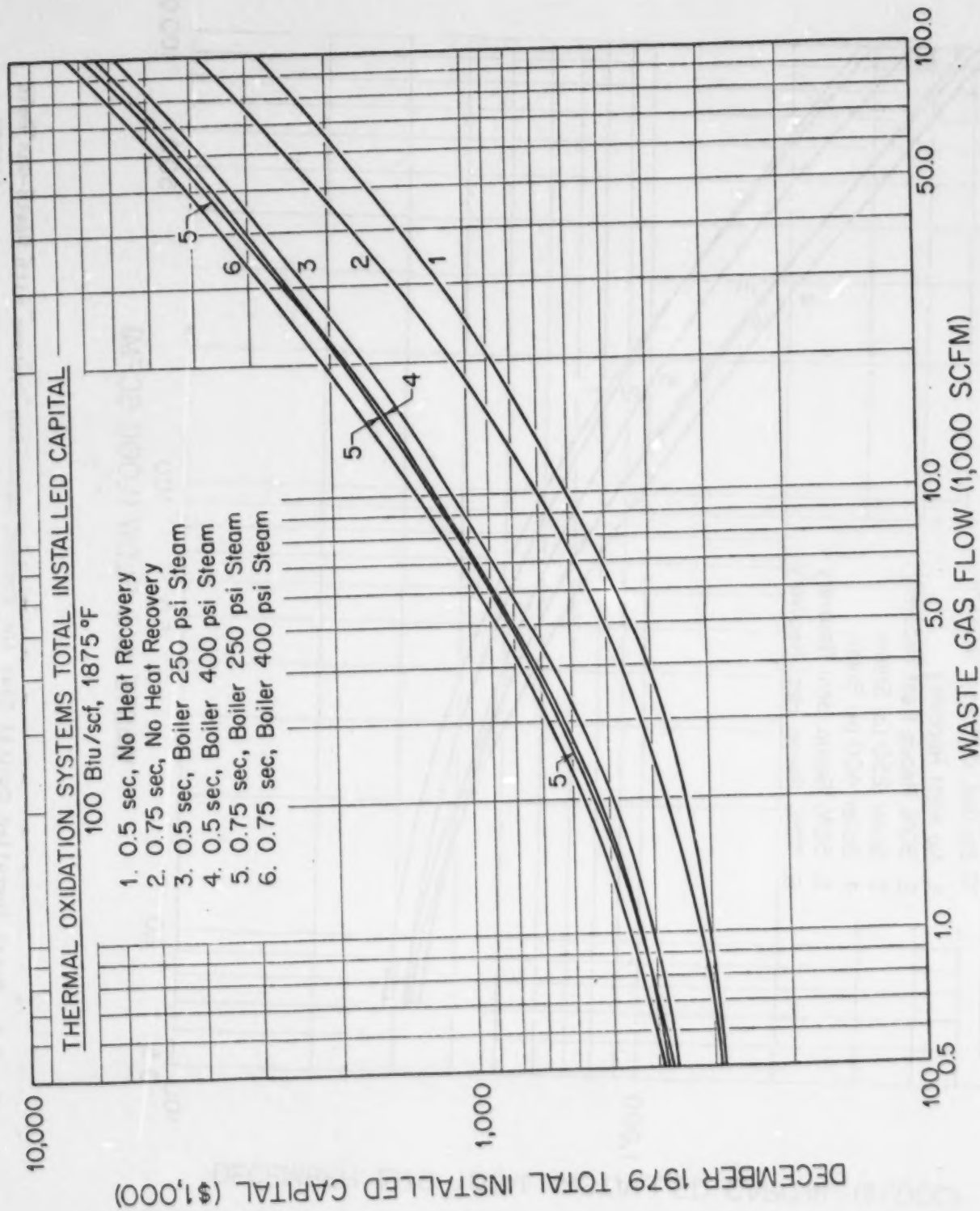


Fig. V-5. Total Installed Capital Cost for Thermal Oxidation Systems with Waste-Gas Heat Content = 100 Btu/scf and Combustion Temperature = 1875°F

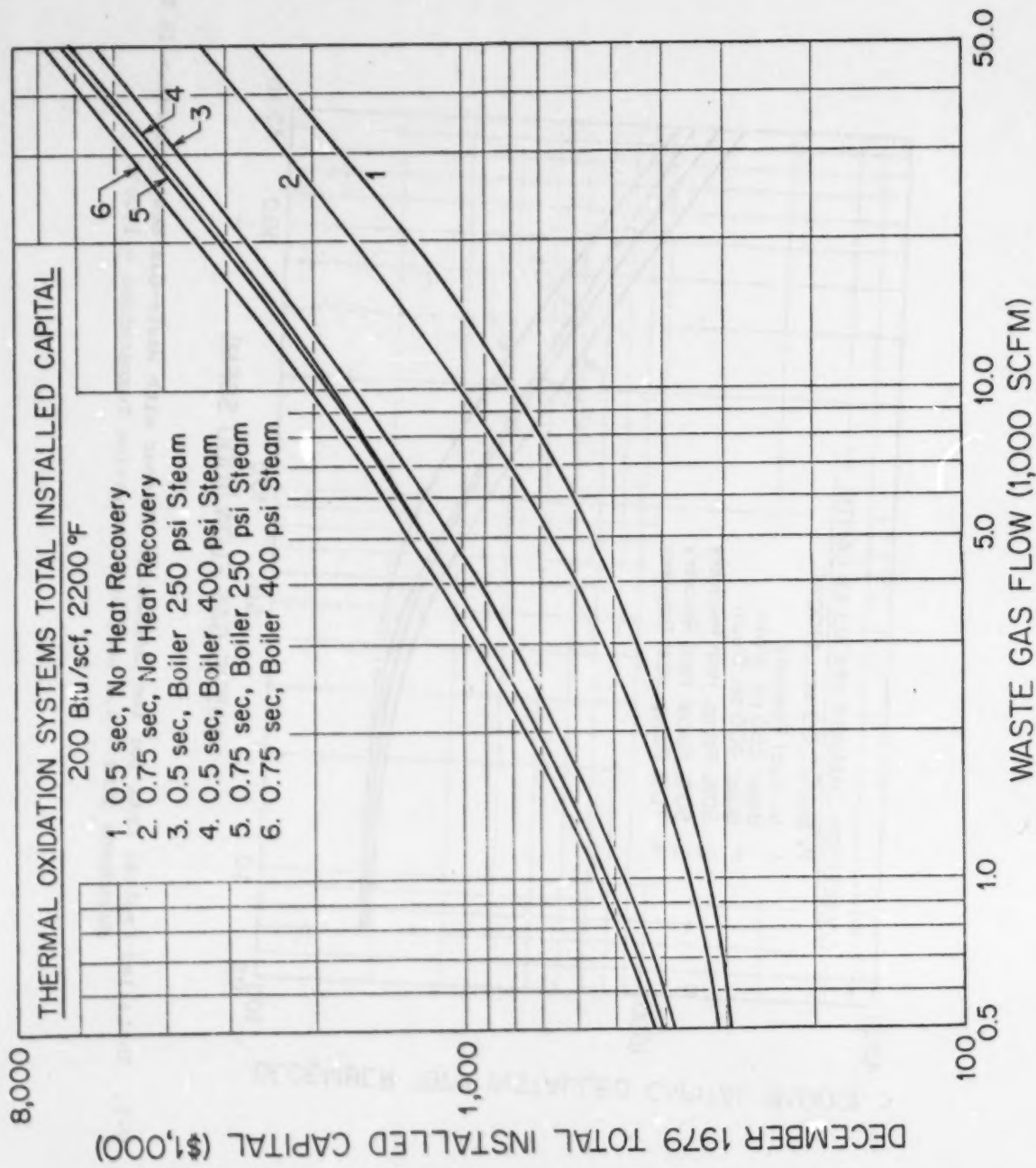


Fig. V-6. Total Installed Capital Cost for Thermal Oxidation Systems with Waste-Gas Heat Content = 200 Btu/scf and Combustion Temperature = 2200 °F

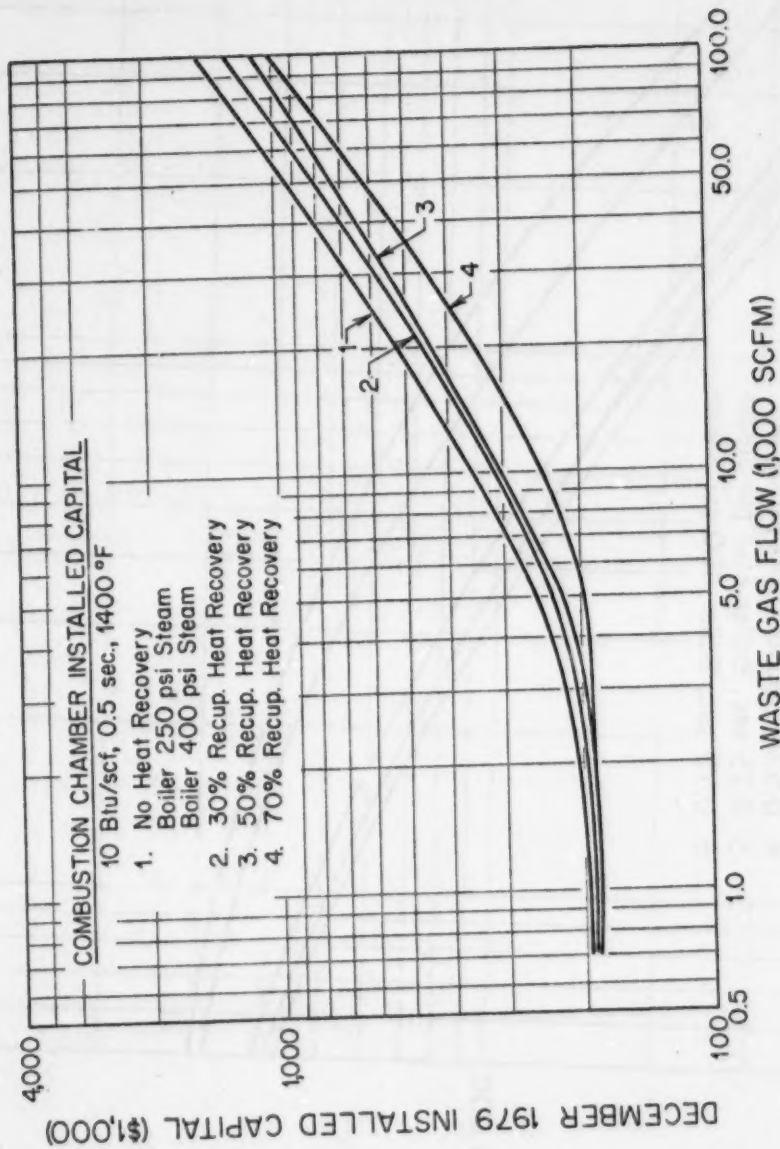


Fig. V-7. Installed Capital Cost for the Combustion Chamber with Waste-Gas Heat Content = 10 Btu/scf, Residence Time = 0.5 sec, and Combustion Temperature = 1400°F

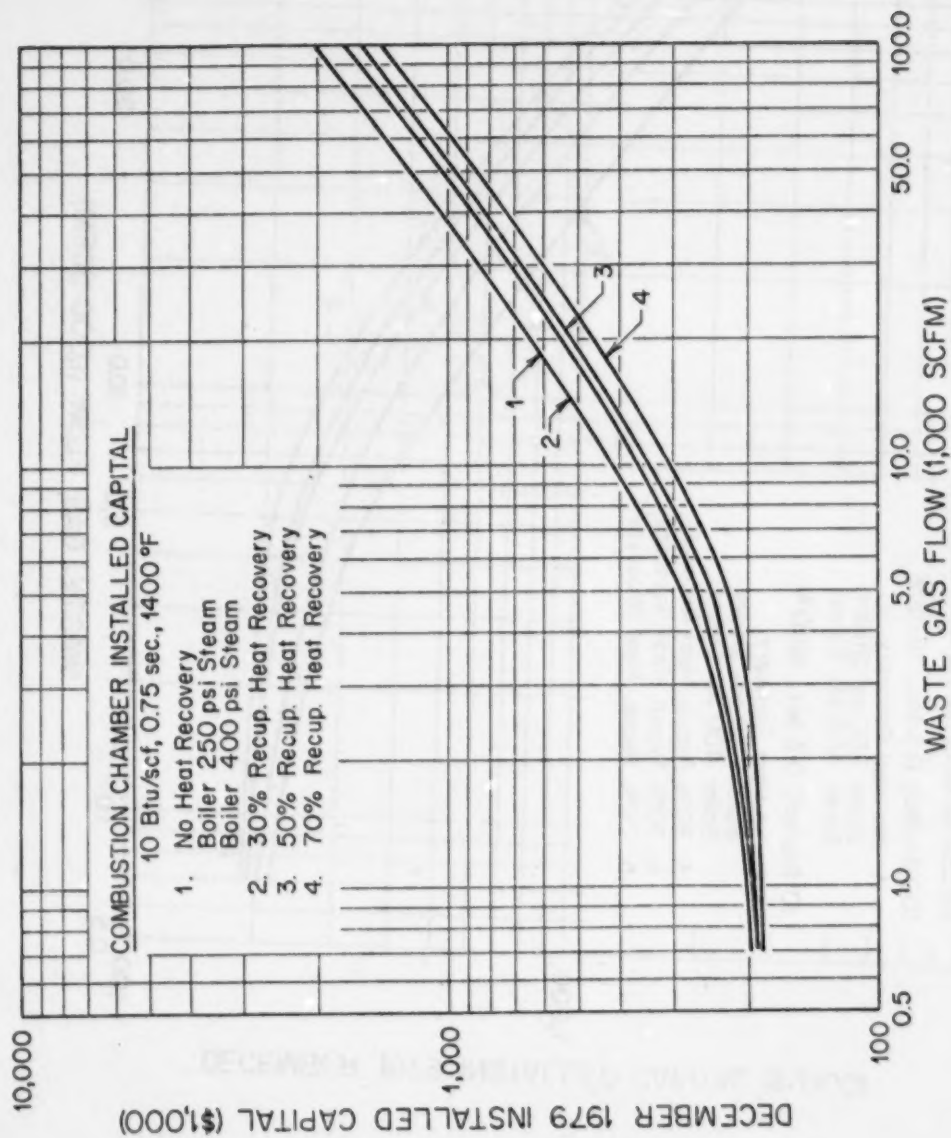


Fig. V-8. Installed Capital Cost for the Combustion Chamber with Waste-Gas Heat Content = 10 Btu/scf, Residence Time = 0.75 sec and Combustion Temperature = 1400°F

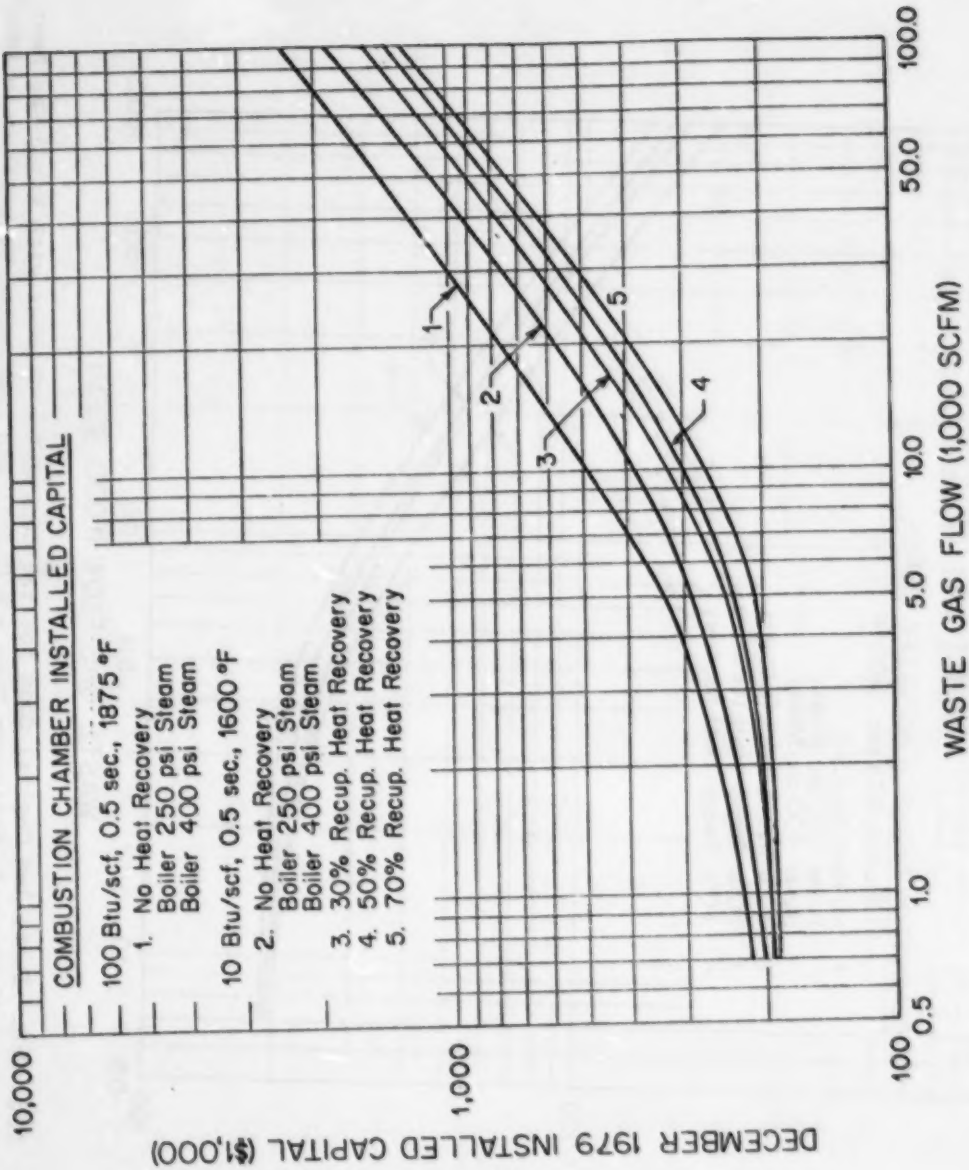


Fig. V-9. Installed Capital Cost for the Combustion Chamber with Waste-Gas Heat Contents = 10 and 100 Btu/scf, Residence Time = 0.5 sec, and Combustion Temperature = 1600 °F

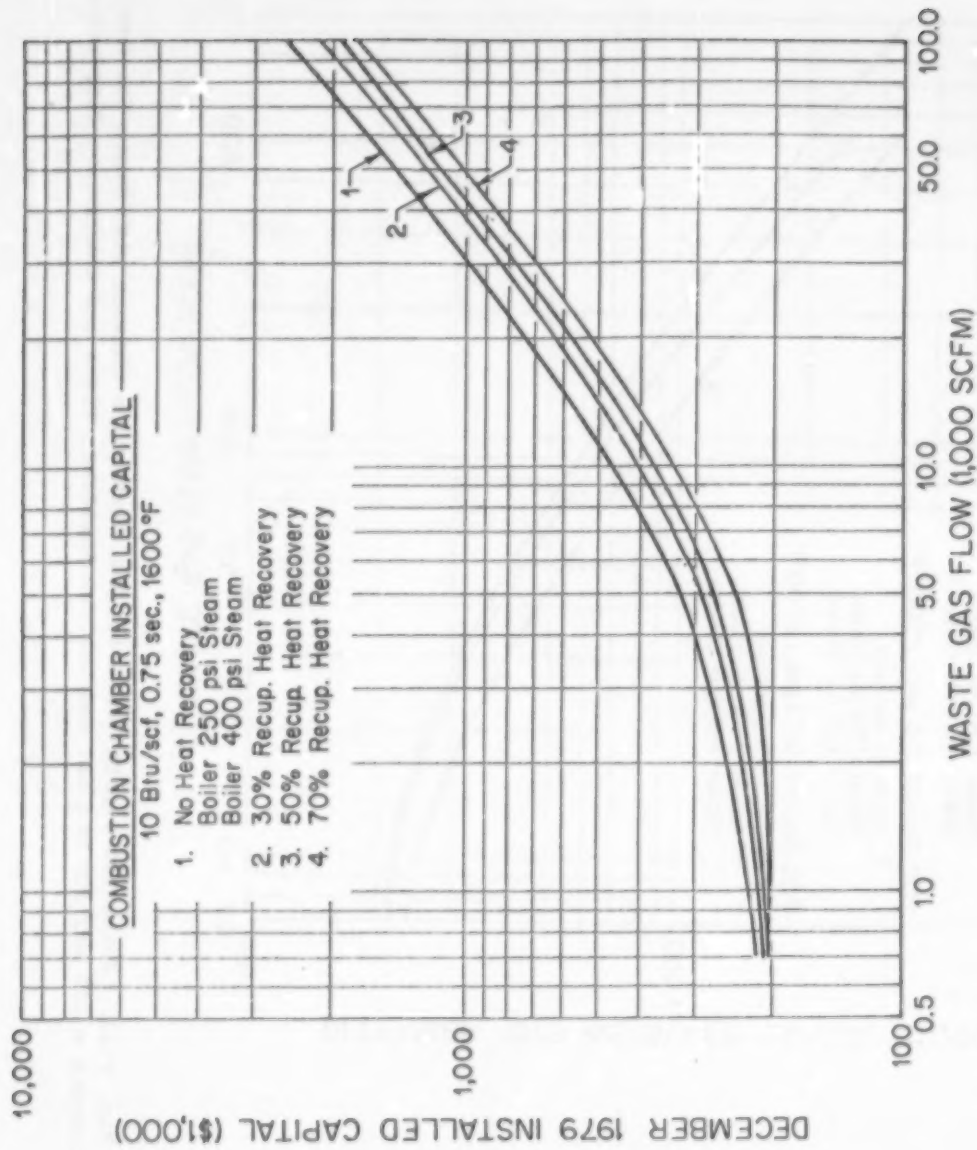


Fig. V-10. Installed Capital Cost for the Combustion Chamber with Waste-Gas Heat Content = 10 Btu/scf, Residence Time = 0.75 sec, and Combustion Temperature = 1600°F

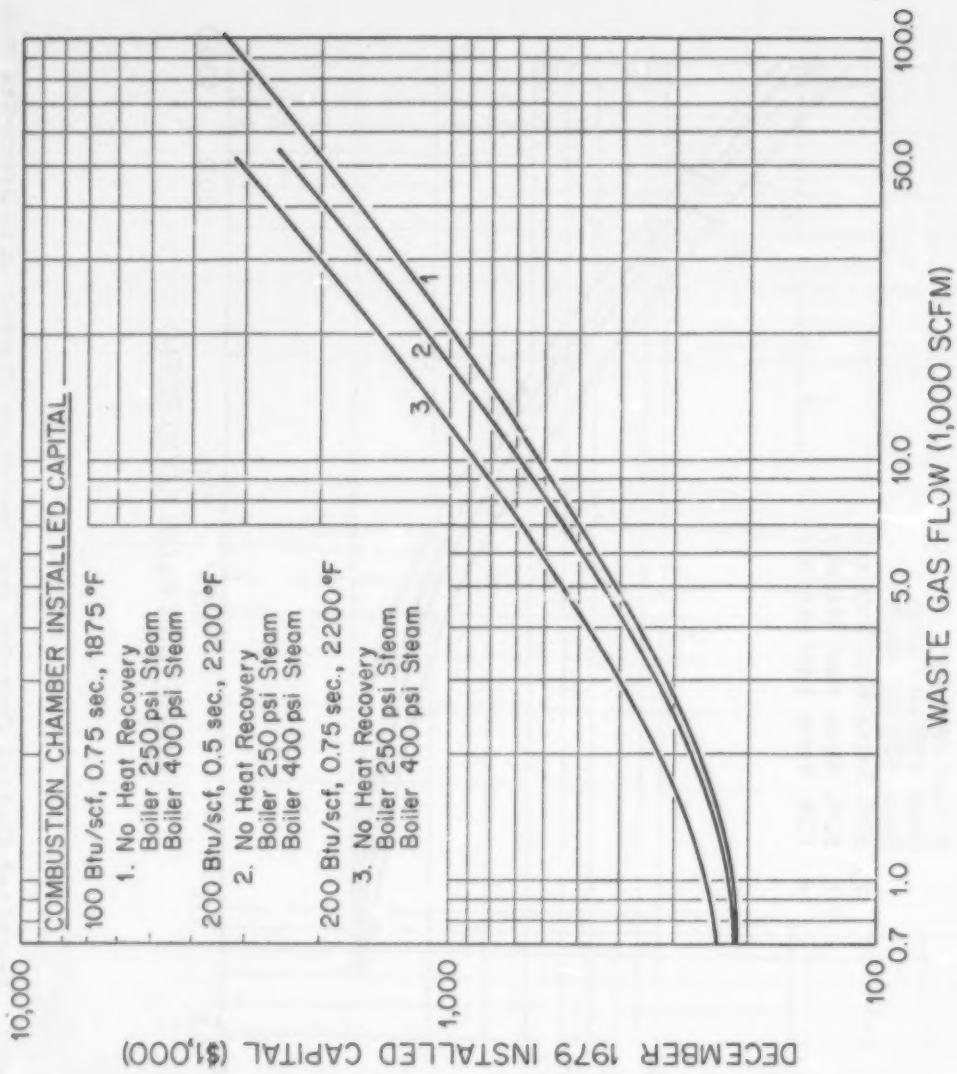


Fig. V-11. Installed Capital Cost for the Combustion Chamber with Waste-Gas Heat Recovery. Contents = 100 and 200 Btu/scf and Various Residence Times and Combustion Temperatures

prepipd and prewired with burners, refractory, etc., all loaded on trucks at the f.o.b. point. Where available and applicable, these vendors also furnished costs for the fan, for heat exchanger/crossover, for the boiler for three steam pressure levels, and for stacks constructed of both refractory and Corten steel. The costs are based on natural-gas auxiliary fuel and include the necessary controls for use of the fuel.

For ease of evaluating system component combinations, all components are presented as a factor of the total waste gas flow in standard cubic feet per minute (scfm).

3. Recuperative Heat Exchangers

Preliminary purchase quotes were obtained from the thermal oxidizer vendors and estimated installed costs were obtained by applying installation factors as previously described. The cost curves of Fig. V-12 for recuperative heat exchangers were developed by the same procedure as that described for combustion chamber cost development.

4. Boiler

Preliminary purchase quotes for steam-generating waste heat boilers were obtained from thermal oxidizer vendors for operation at various steam pressure levels and various proportions of heat recovery. The estimated installed costs were then obtained by applying installation factors to these purchase costs as previously described. The cost curves, shown in Figs. V-13 and 14 for the boilers, were developed as described previously.

5. Fans, Ductwork, and Stacks

The installed capital costs for the fans, ducts, and stacks are plotted in Figs. V-15, 16, and 17 for systems with no heat recovery, with recuperative heat recovery, and with waste-heat boilers respectively.

- a. Ducts—Each system is assumed to require 150 ft of round-steel inlet ductwork with four ells, one expansion joint, and one damper with an actuator. The cost data source is a report by GARD, Inc.,¹⁵ which was prepared for the EPA and includes total installed costs.

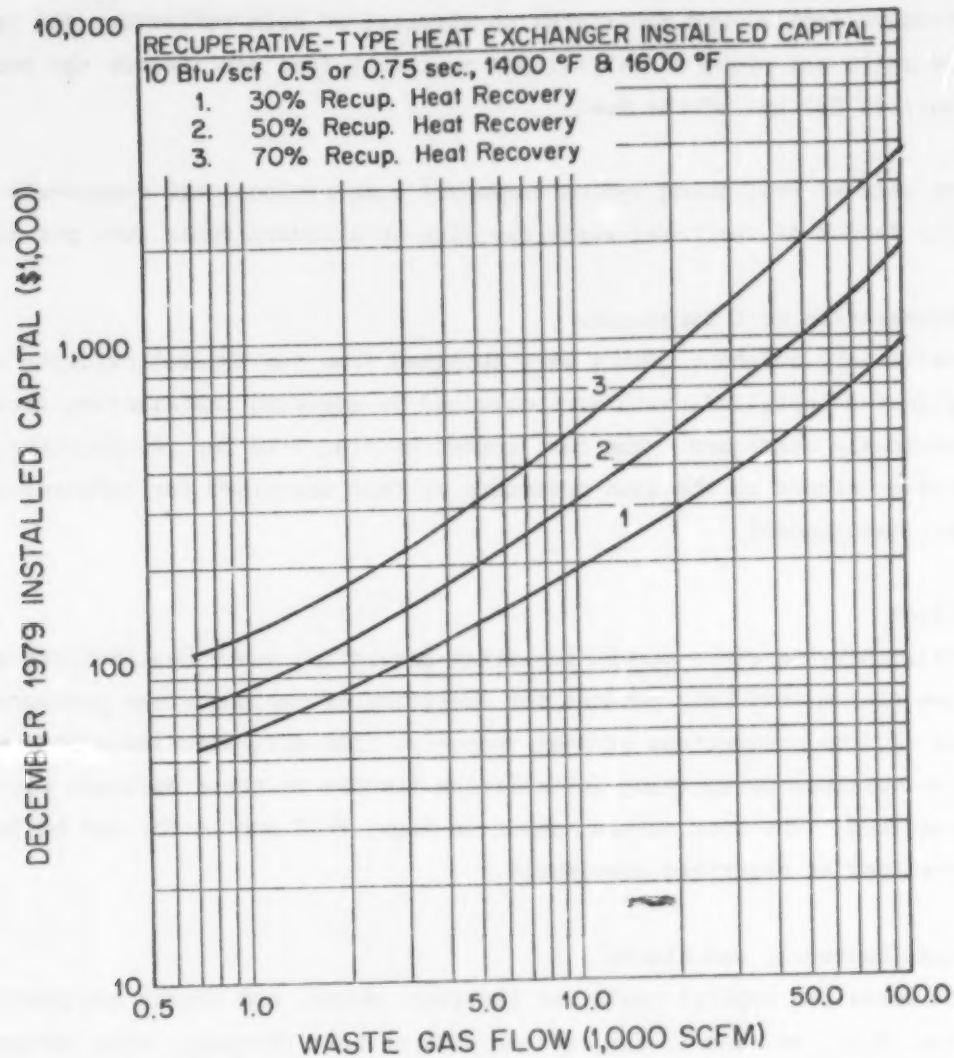


Fig. V-12. Installed Capital Cost for Recuperative-Type Heat Exchangers with the Waste-Gas Heat Content = 10 Btu/scf

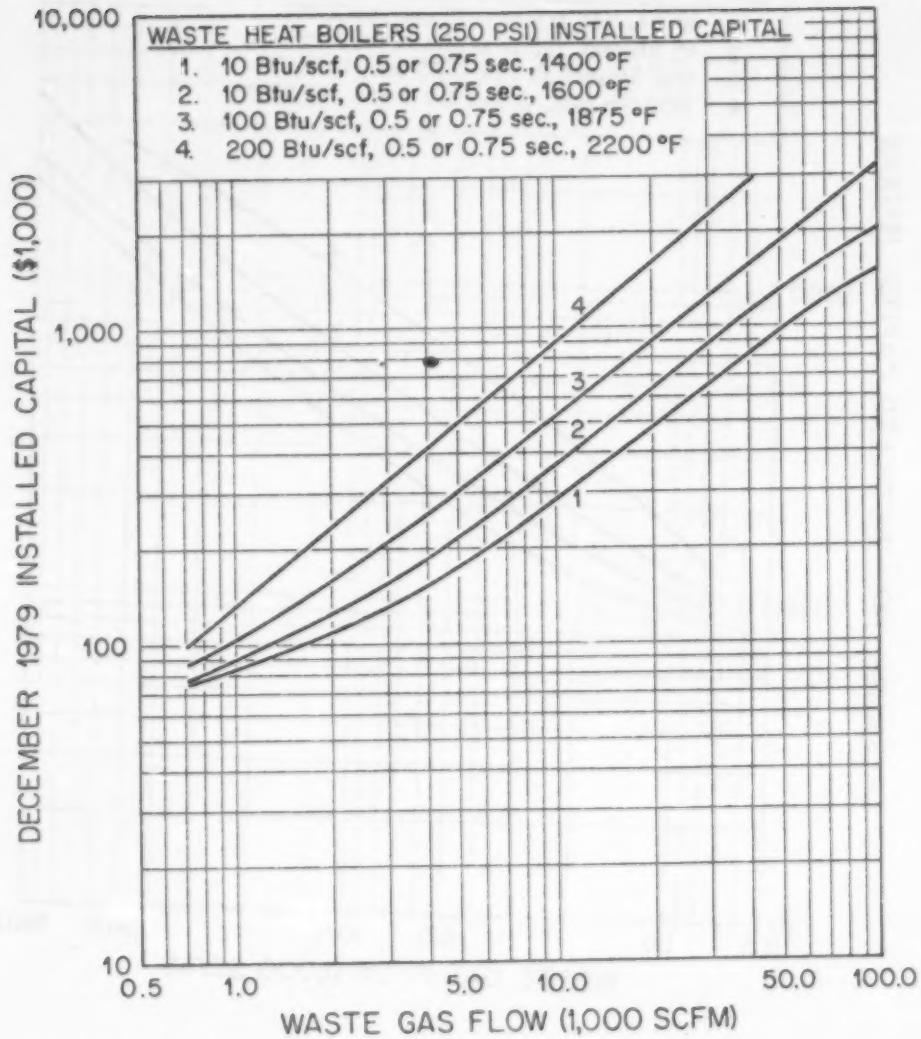


Fig. V-13. Installed Capital Cost for Waste Heat Boilers (250 psi)

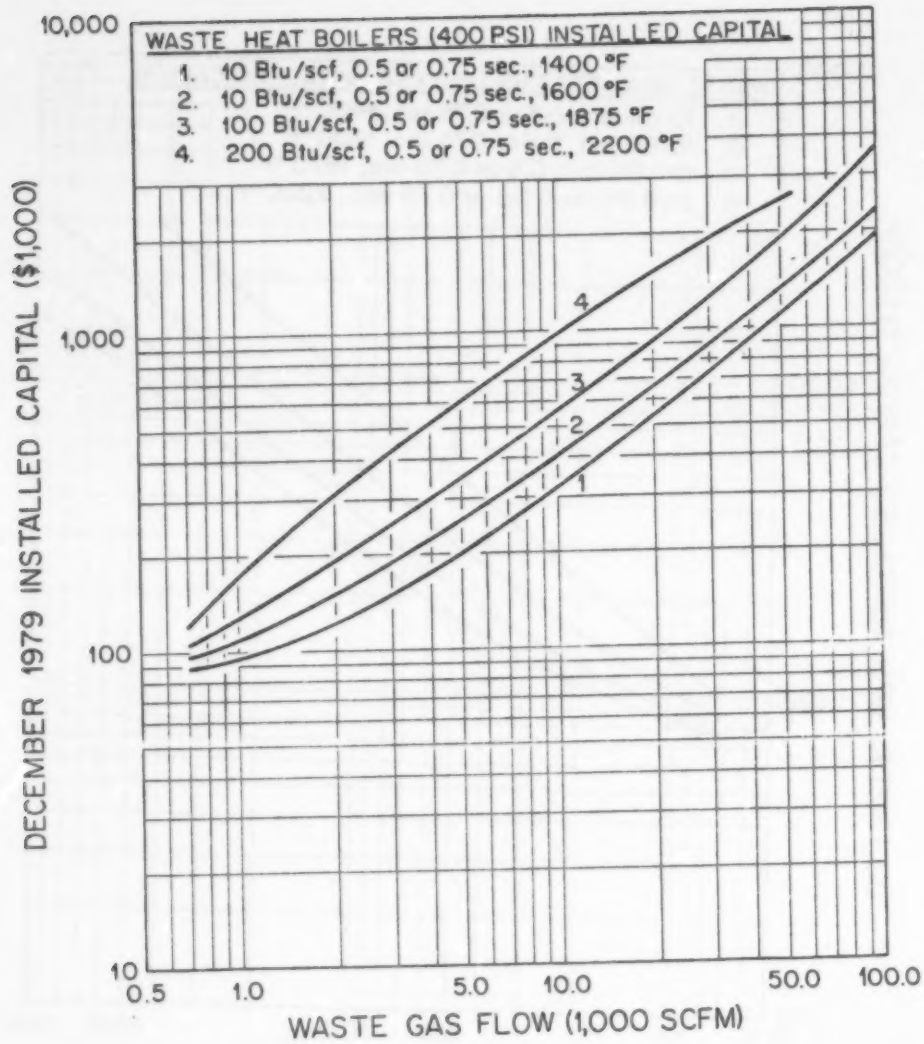


Fig. V-14. Installed Capital Cost for Waste Heat Boilers (400 psi)

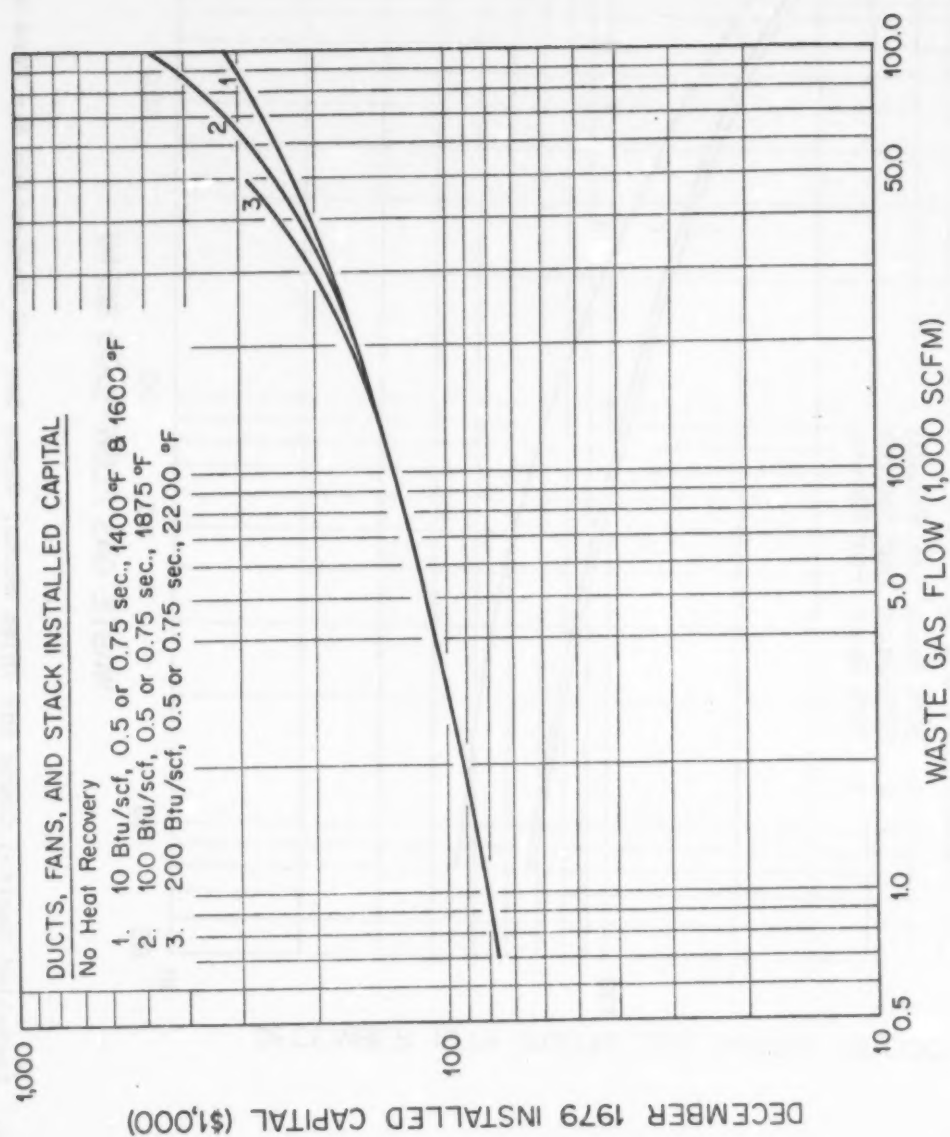


Fig. V-15. Installed Capital Costs for Inlet Ducts, Waste Gas, and Combustion Air Fans and Stack for System with No Heat Recovery

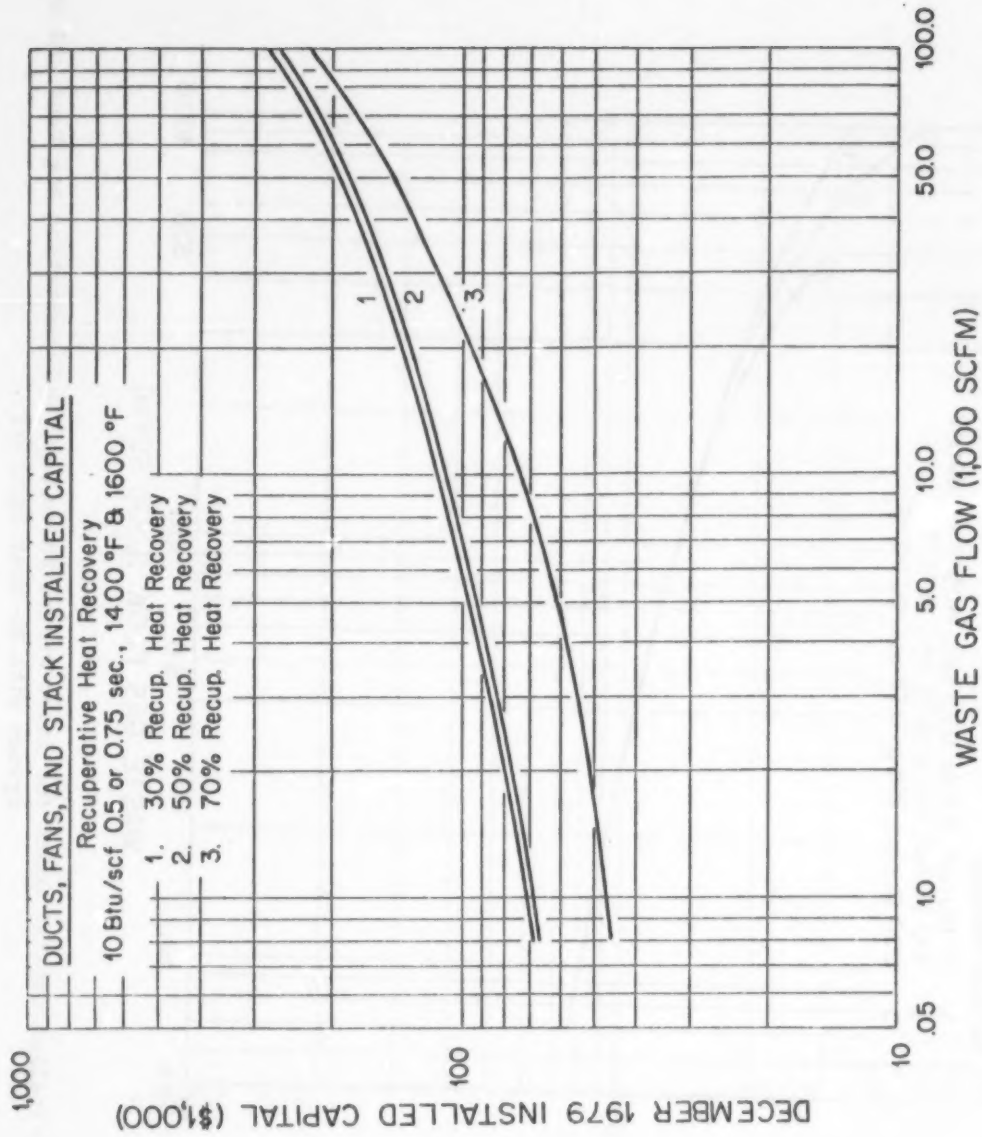


Fig. V-16. Installed Capital Costs for Inlet Ducts, Waste Gas and Combustion Air Fans and Stack with Recuperative Heat Recovery

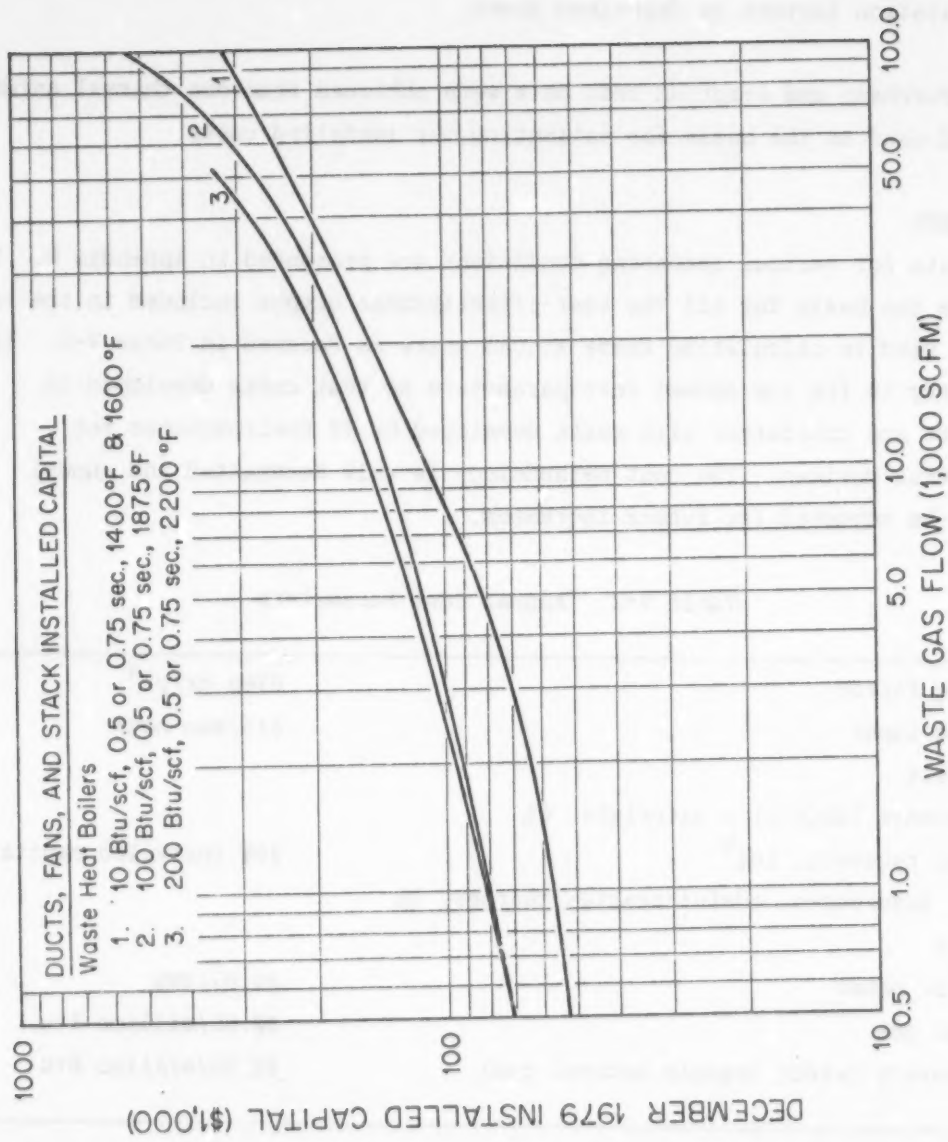


Fig. V-17. Installed Capital for Inlet Ducts, Waste Gas, and Combustion Air Fans and Stack with Waste Heat Boilers

- b. Fans—Each system is assumed to require both a process-waste-gas and combustion air fan. Various pressure-head requirements were assumed to match system requirements, such as increased pressure drop caused by incorporation of the heat recovery device. The cost data source is a tabulation of equipment purchase costs by Richardson Engineering Co.¹⁶ The installed costs shown were obtained by applying installation factors as described above.
- c. Stacks—Purchase and erection cost data were obtained from one thermal oxidizer vendor and used as the basis for estimating the installed costs.
- C. ANNUAL COSTS
- Annual costs for various operating conditions are presented in Appendix B. These costs were the basis for all the cost-effectiveness graphs included in the report. The basis used in calculating these annual costs is defined in Table V-2. It is necessary to fix the annual cost parameters so that costs developed in this report are consistent with costs developed by IT Enviroscience for other control devices. The cost methodology is well documented and annual costs can be adjusted for future increases.

Table V-2. Annual Cost Parameters

Operating factor	8760 hr/yr ^a
Operating labor	\$15/man-hour
Fixed costs	
Maintenance labor plus materials, 6%	
Capital recovery, 18% ^b	29% installed capital
Taxes, insurances, administrative charges, 5%	
Utilities	
Electric power	\$0.03/kWh
Natural gas	\$2.00/million Btu
Heat recovery credit (equals natural gas)	\$2.00/million Btu

^a Process downtime is normally expected to range from 5 to 15%. If the hourly rate remains constant, the annual production and annual VOC emissions will be correspondingly reduced. Control devices will usually operate on the same cycle as the process. From the standpoint of cost-effectiveness calculations, the error introduced by assuming continuous operation is negligible.

^b Based on 10-year life and 12% interest.

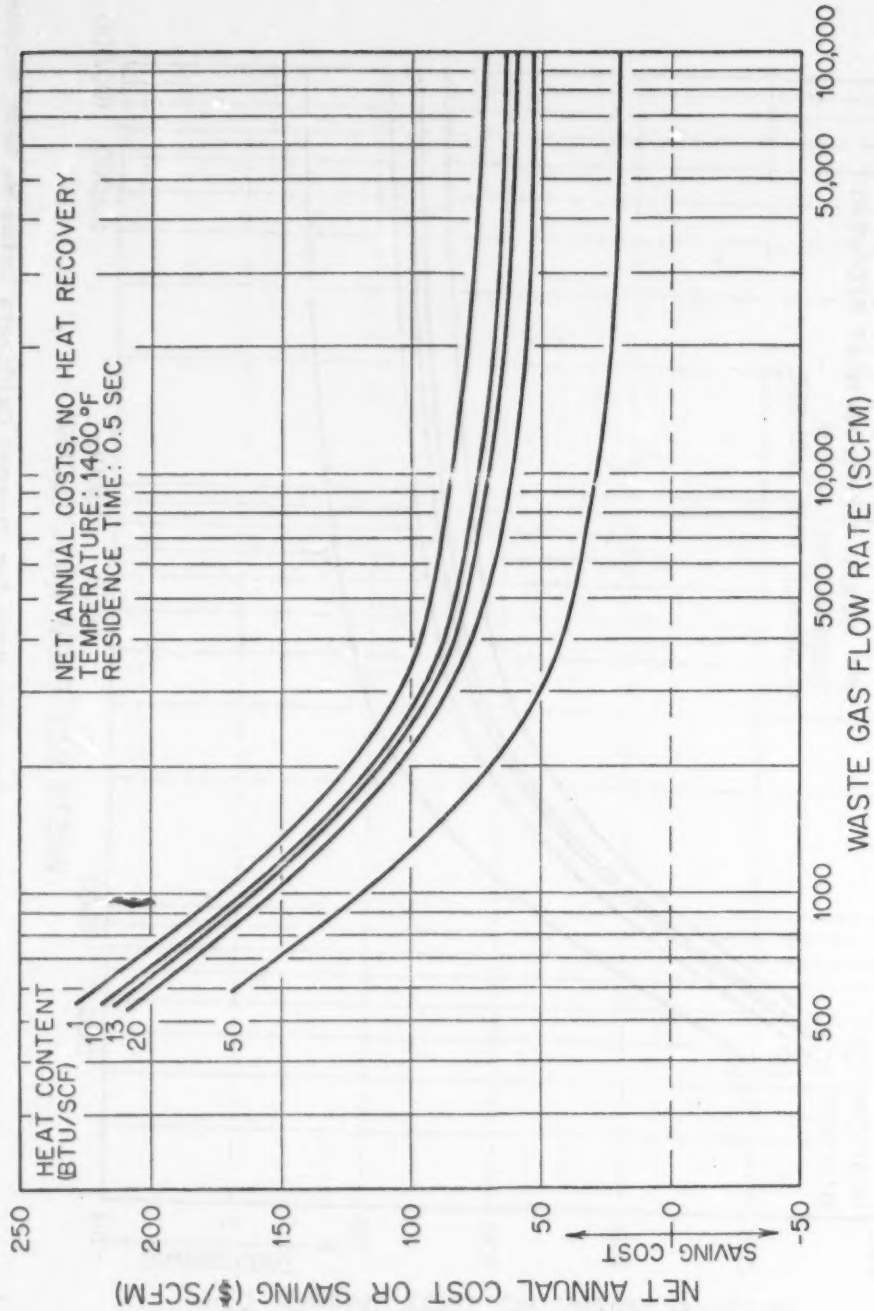


Fig. V-18. Net Annual Costs vs Waste Gas Flow Rate for Thermal Oxidizers Using No Heat Recovery, 1400°F Combustion Temperature, 0.5 sec Residence Time, and Heat Contents from 1 to 50 Btu/scf

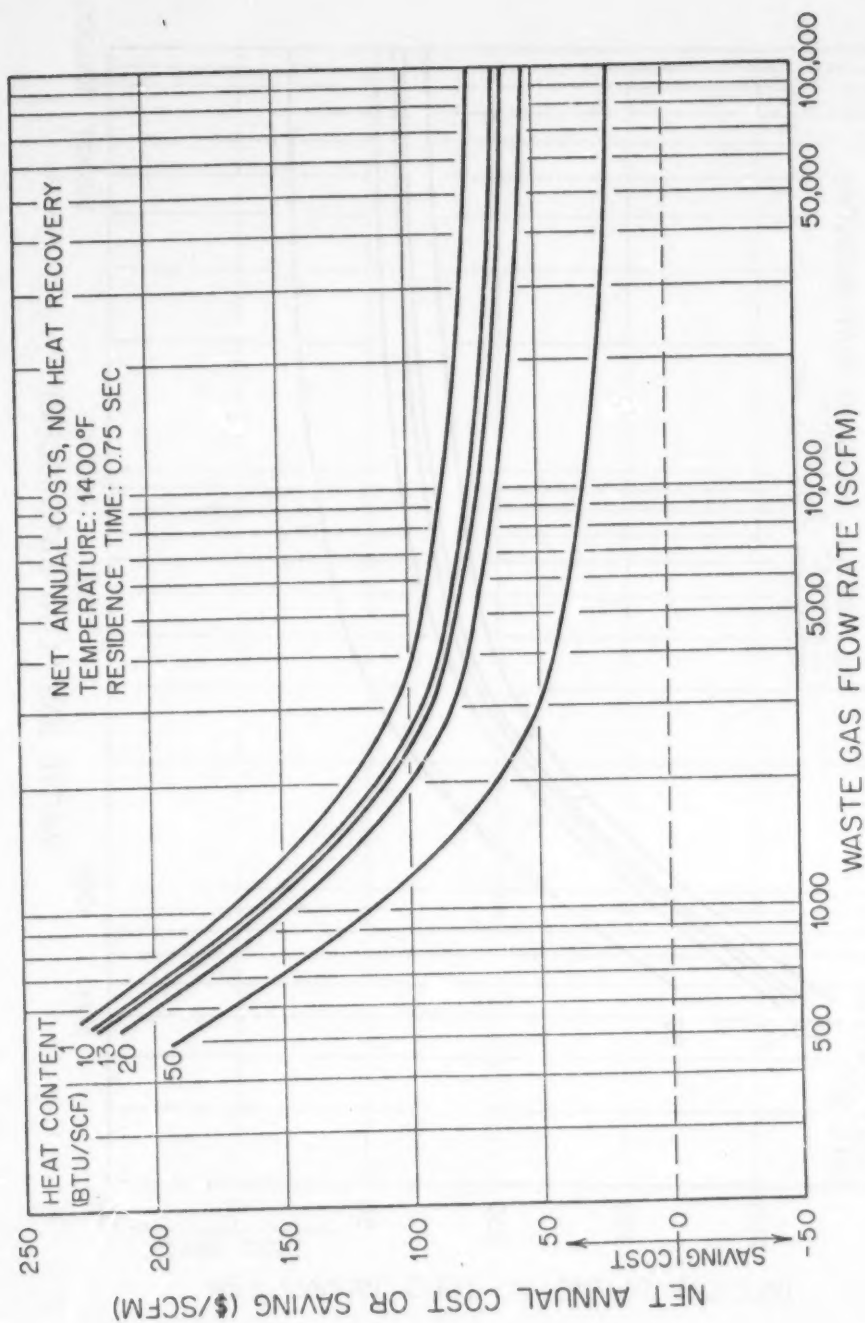


Fig. V-19. Net Annual Costs vs Waste Gas Flow Rate for Thermal Oxidizers Using No Heat Recovery, 1400°F Combustion Temperature, 0.75 sec Residence Time, and Heat Contents from 1 to 50 Btu/scf

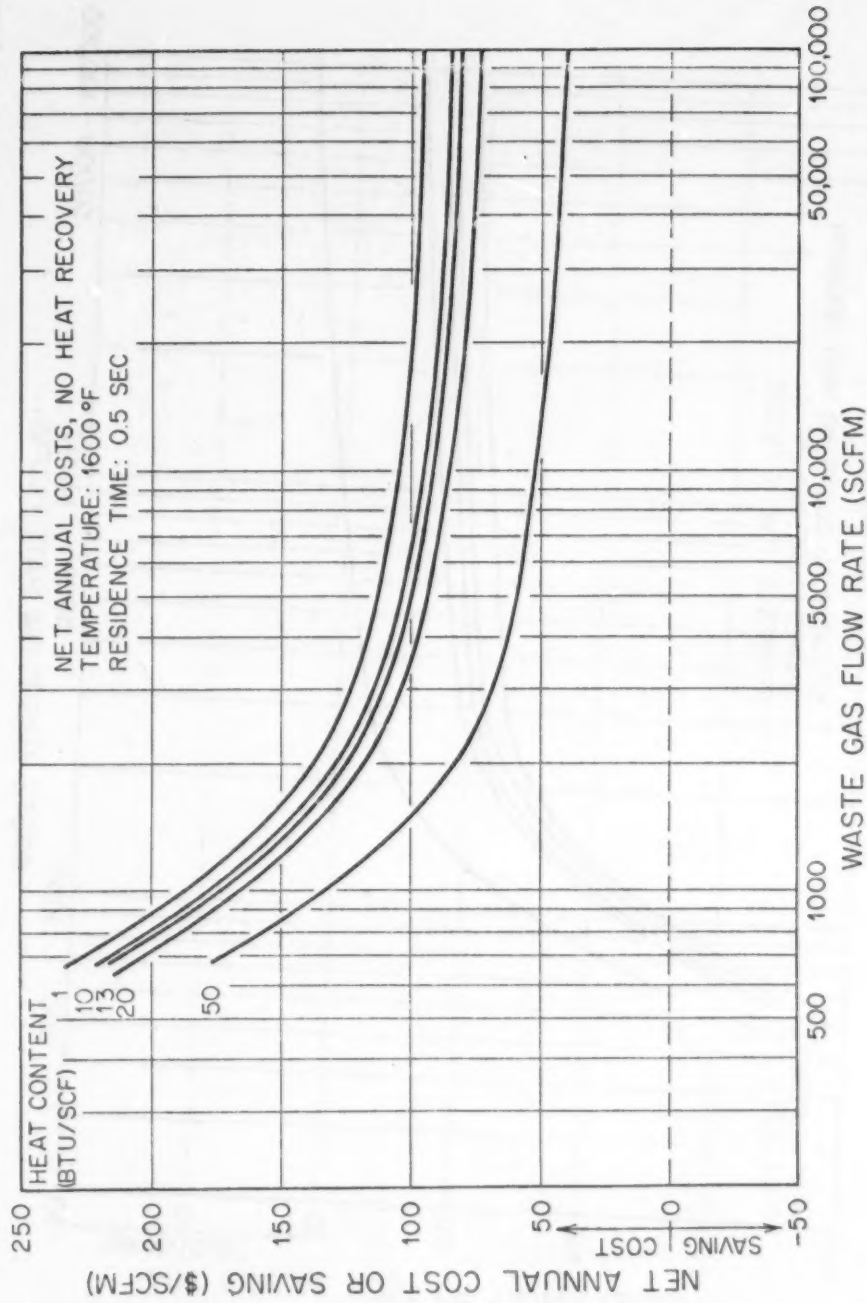


Fig. V-20. Net Annual Costs vs Waste Gas Flow Rate for Thermal Oxidizers Using No Heat Recovery, 1600°F Combustion Temperature, 0.5 sec Residence Time, and Heat Contents from 1 to 50 Btu/scf

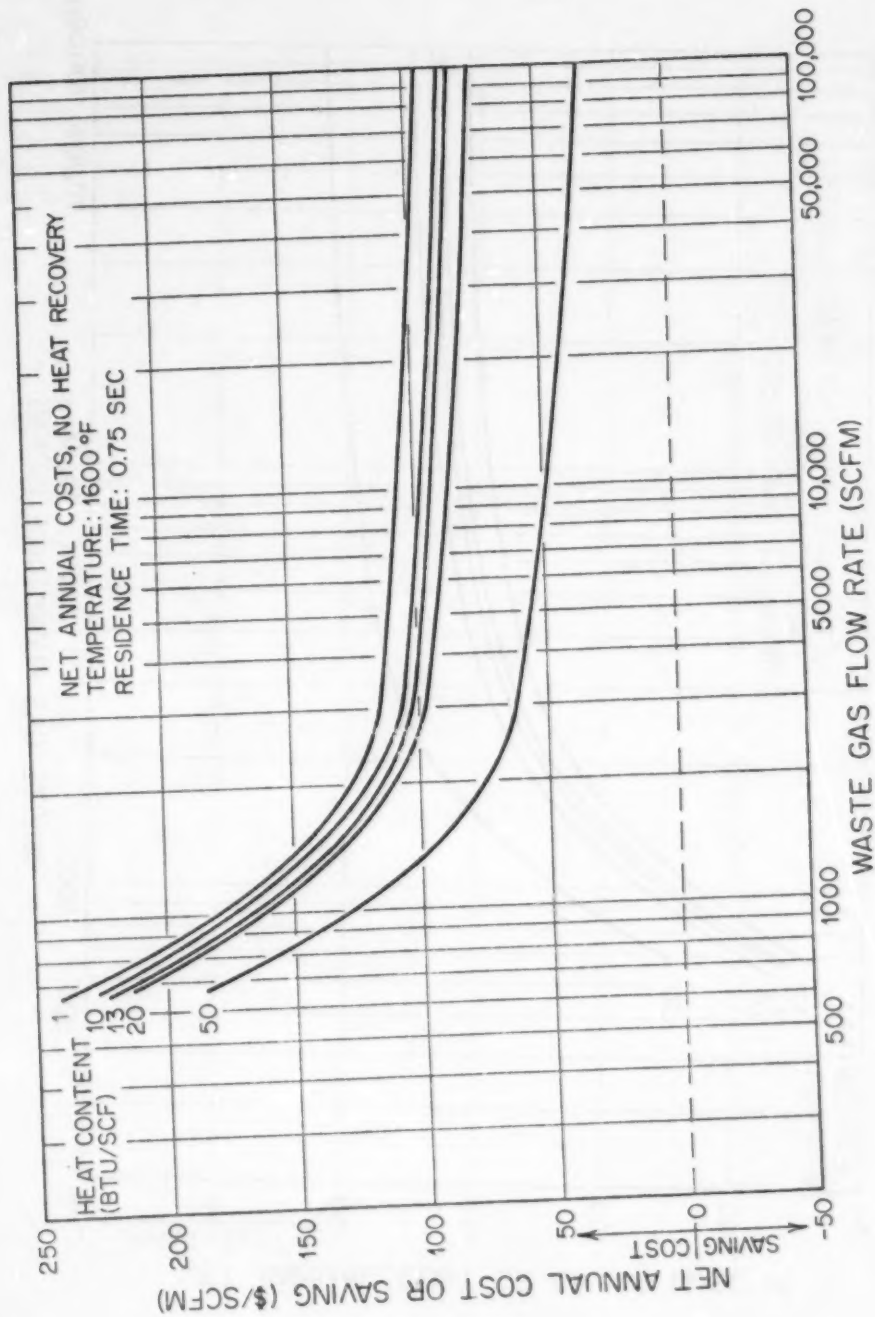


Fig. V-21. Net Annual Costs vs Waste Gas Flow Rate for Thermal Oxidizers Using No Heat Recovery, 1600°F Combustion Temperature, 0.75 sec Residence Time, and Heat Contents from 1 to 50 Btu/scf

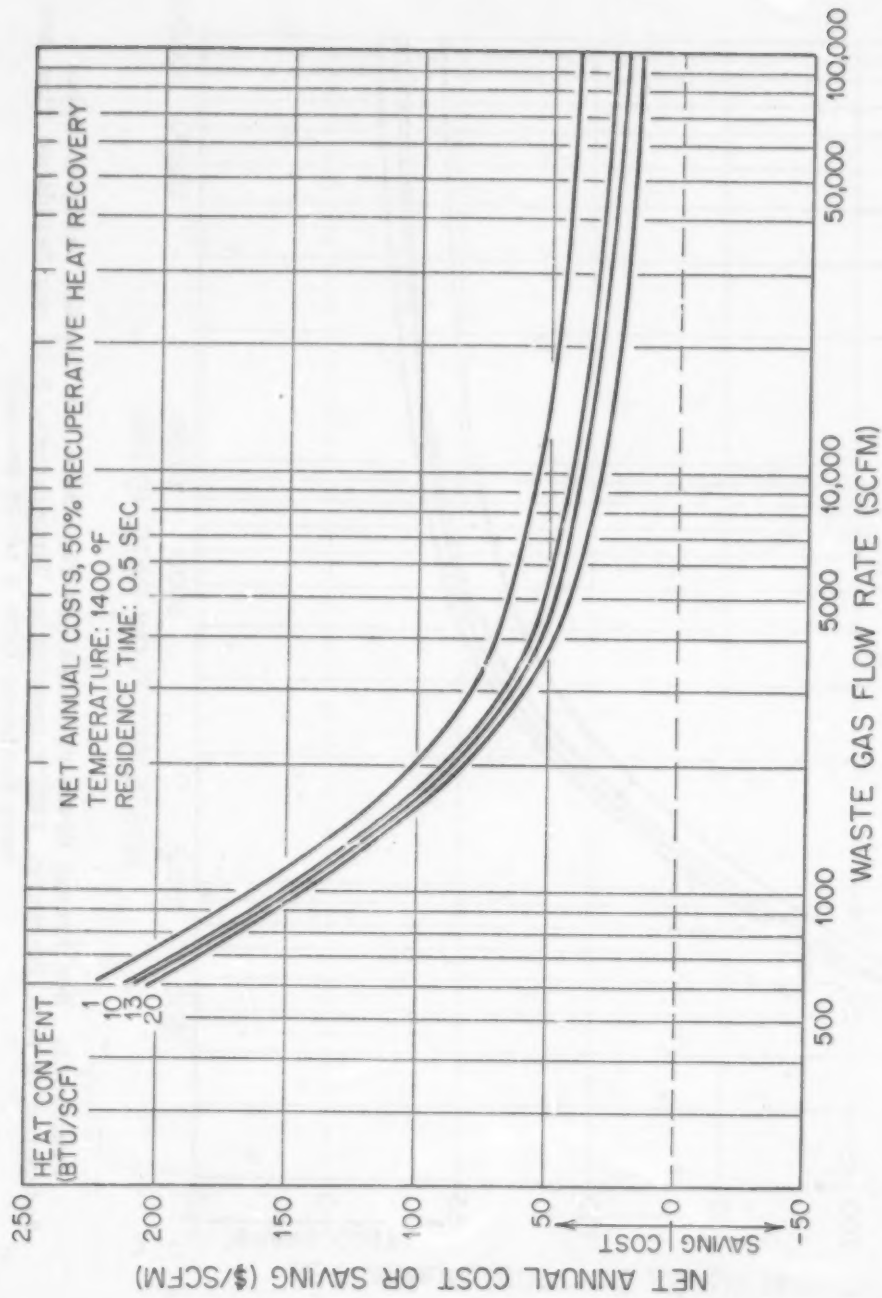


Fig. V-22. Net Annual Costs vs Waste Gas Flow Rate for Thermal Oxidizers Using Recuperative Heat Recovery, 1400°F Combustion Temperature, 0.5 sec Residence Time, and Heat Contents from 1 to 20 Btu/scf

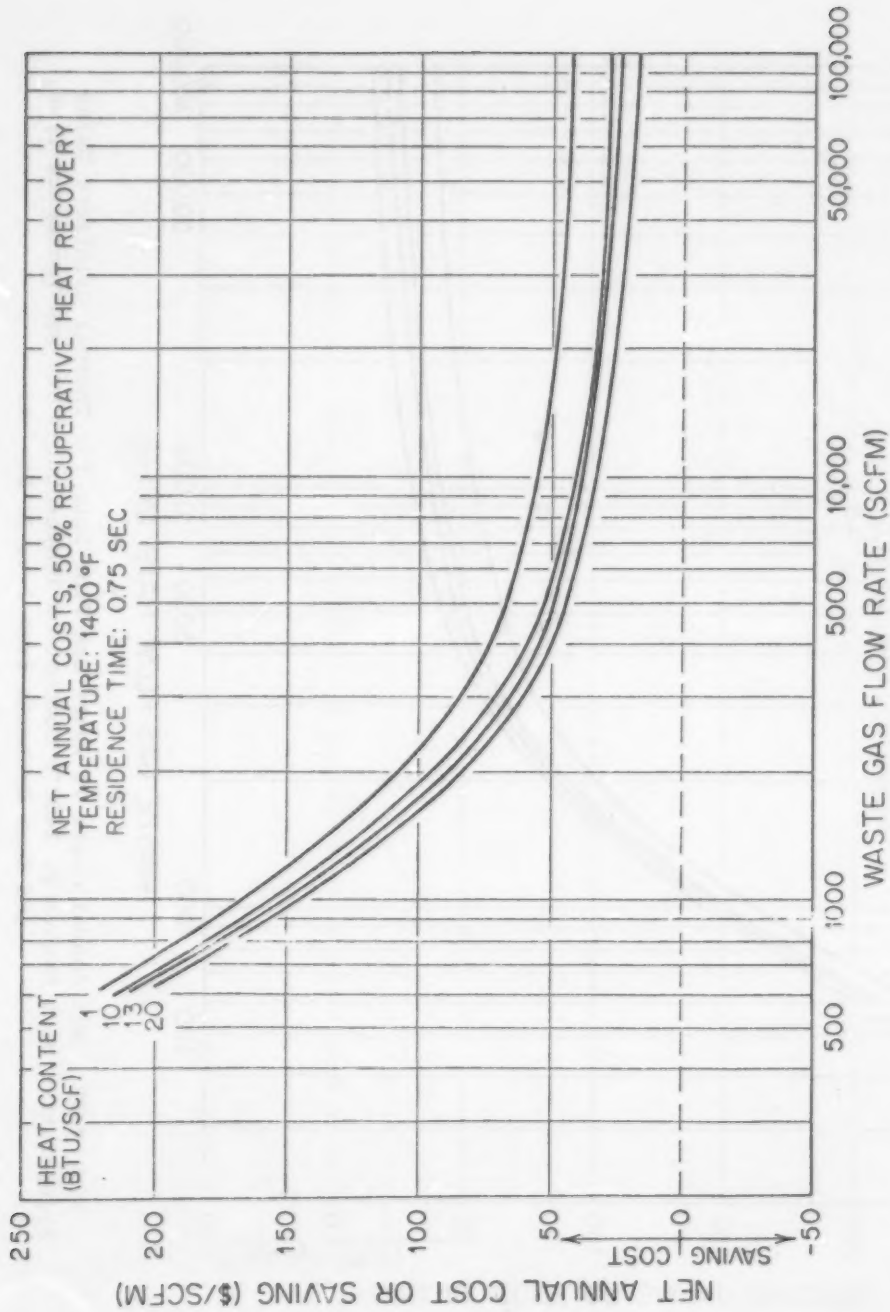


Fig. V-23. Net Annual Costs vs Waste Gas Flow Rate for Thermal Oxidizer Using Recuperative Heat Recovery, 1400°F Combustion Temperature, 0.75 sec Residence Time and Heat Contents from 1 to 20 Btu/scf

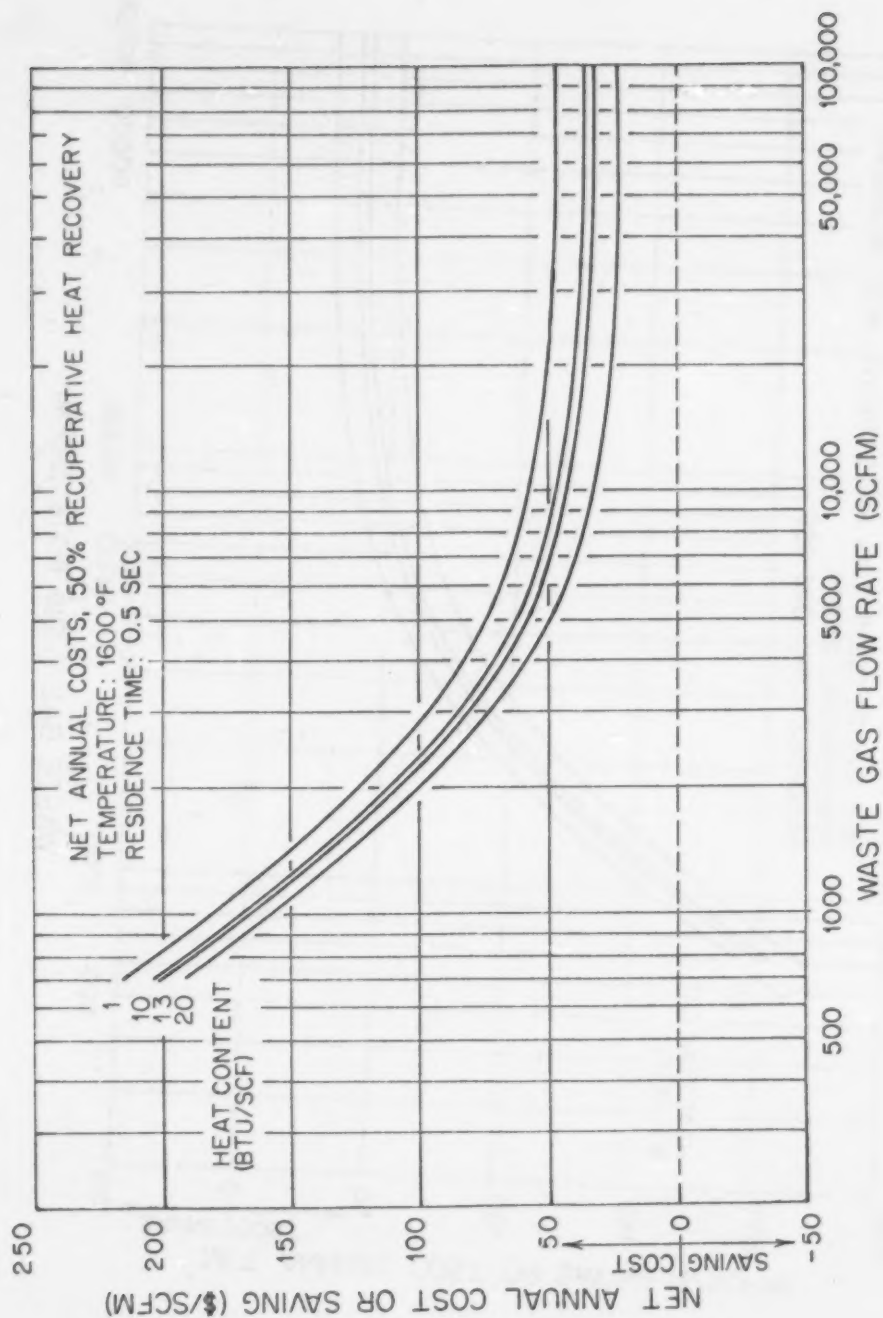


Fig. V-24. Net Annual Cost vs Waste Gas Flow Rate for Thermal Oxidizer Using Recuperative Heat Recovery, 1600°F Combustion Temperature, 0.5 sec Residence Time and Heat Contents from 1 to 20 Btu/scf

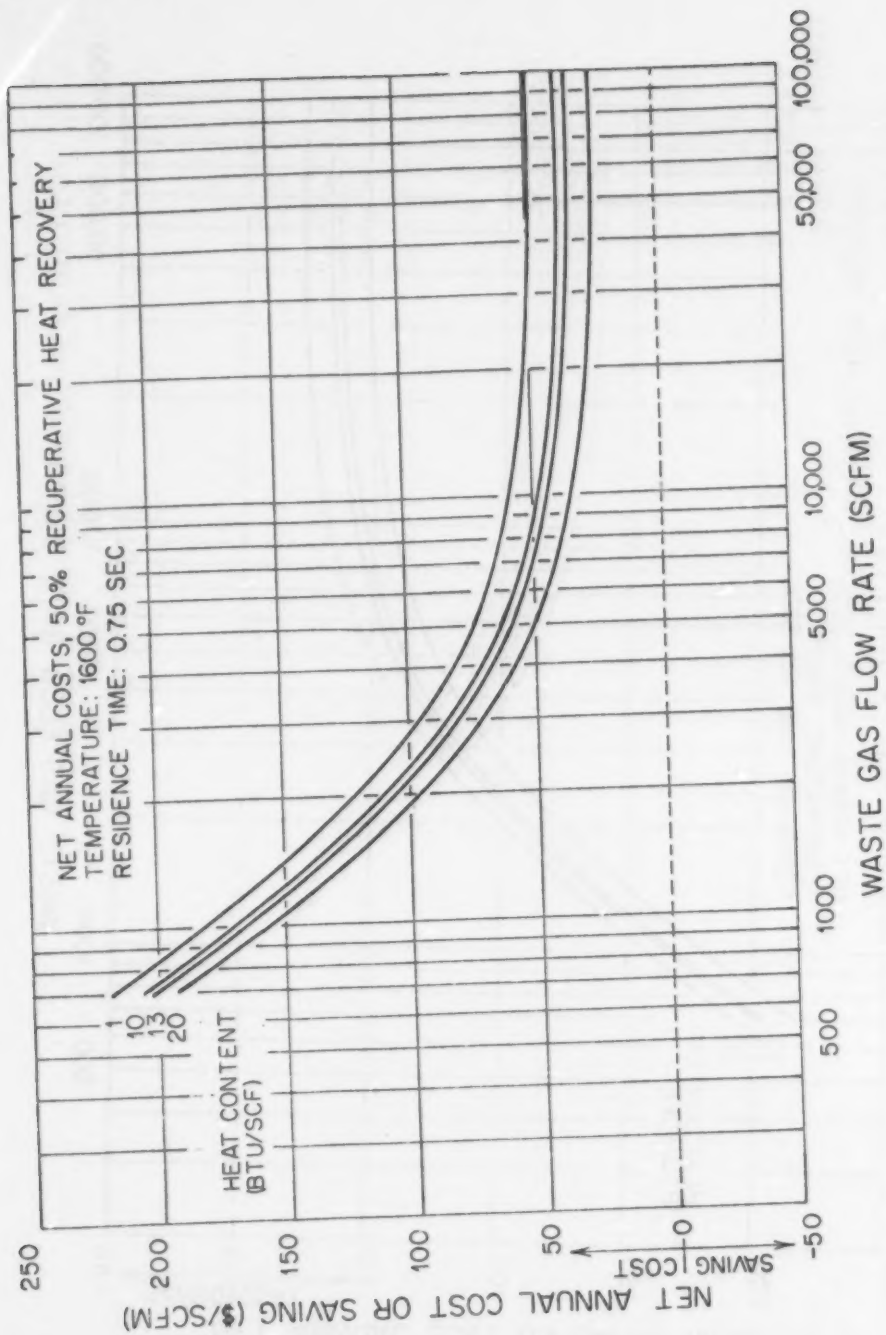


Fig. V-25. Net Annual Cost vs Waste Gas Flow Rate for Thermal Oxidizer Using Recuperative Heat Recovery, 1600°F Combustion Temperature, 0.75 sec Residence Time and Heat Contents from 1 to 20 Btu/scf

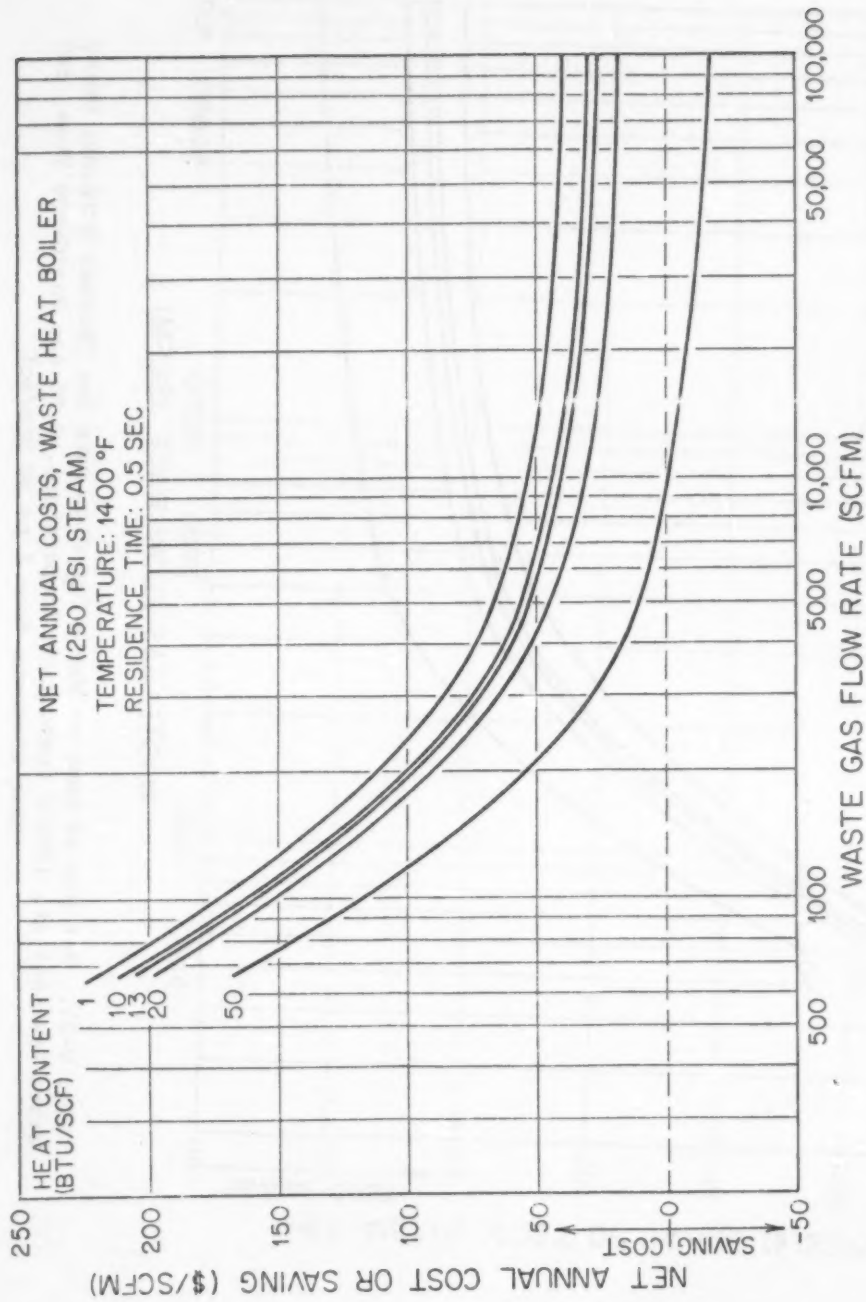


Fig. V-26. Net Annual Cost vs Waste Gas Flow Rate for Thermal Oxidizer Using Waste Heat Boiler, 1400°F Combustion Temperature, 0.5 sec Residence Time, and Heat Contents from 1 to 50 Btu/scf

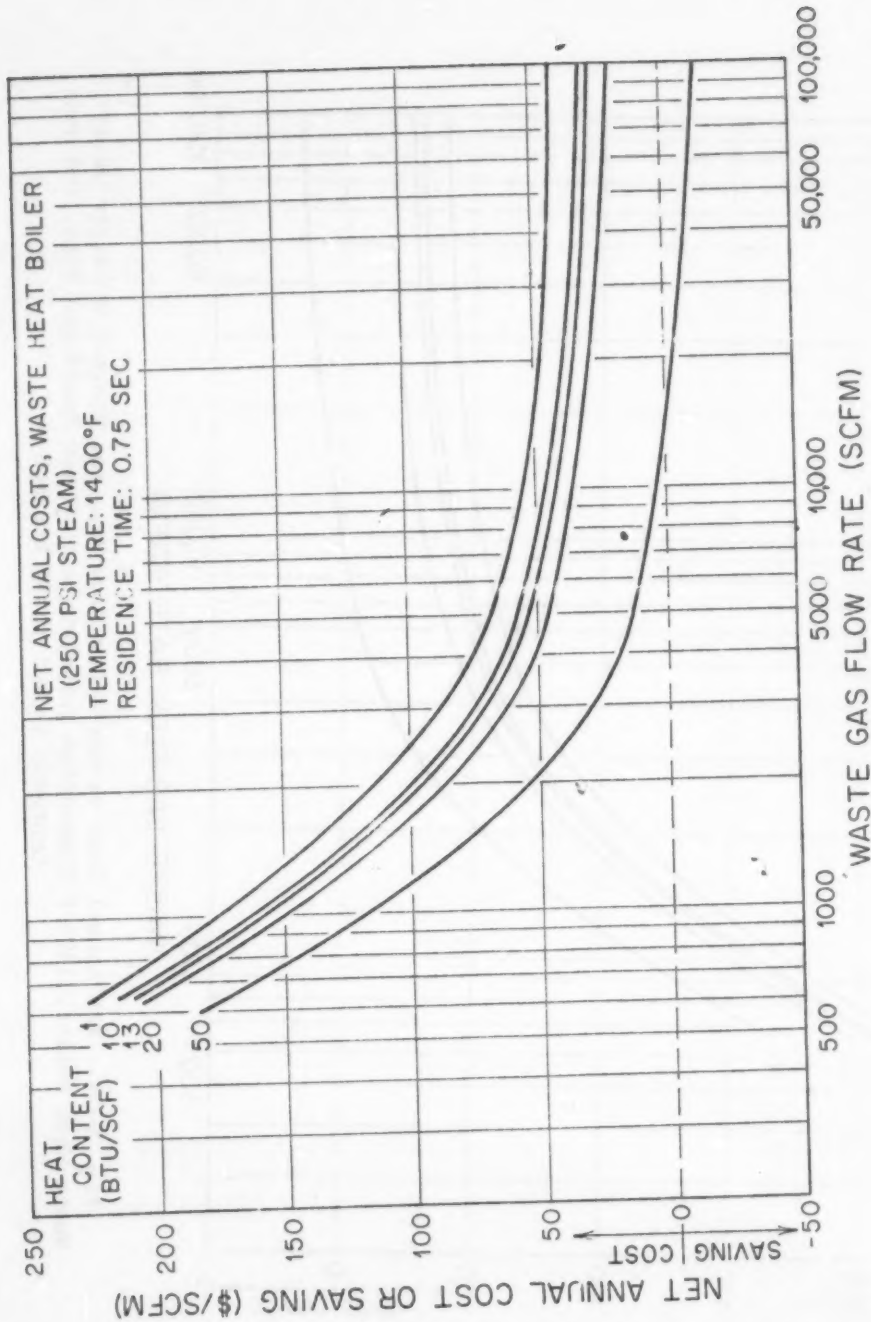


Fig. V-27. Net Annual Cost vs Waste Gas Flow Rate for Thermal Oxidizer Using Waste Heat Boiler, 1400°F Combustion Temperature, 0.75 sec Residence Time, and Heat Contents from 1 to 50 Btu/scf

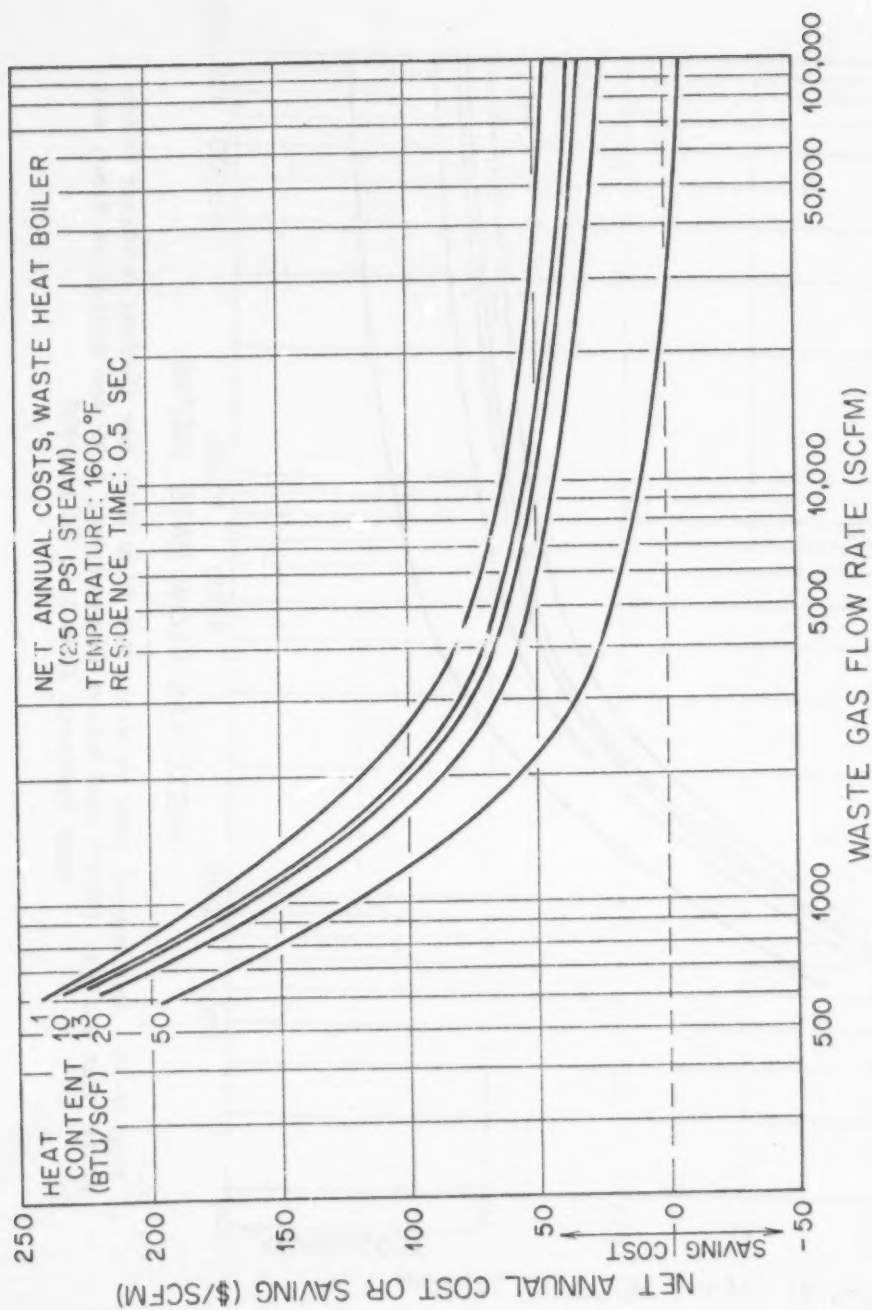


Fig. V-28. Net Annual Cost vs Waste Gas Flow Rate for Thermal Oxidizer Using Waste Heat Boiler, 1600°F Combustion Temperature, 0.5 sec Residence Time, and Heat Contents from 1 to 50 Btu/scf

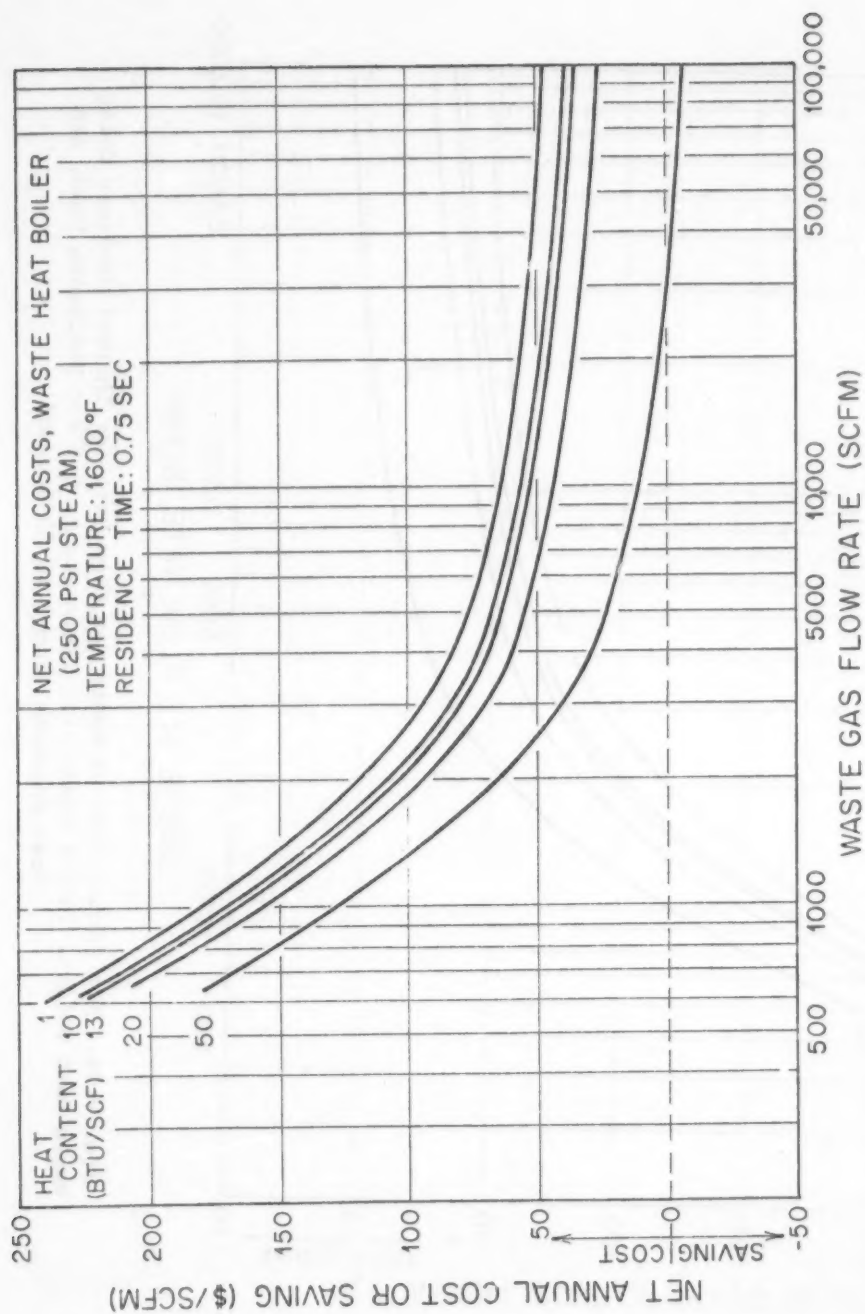


Fig. V-29. Net Annual Cost vs Waste Gas Flow Rate for Thermal Oxidizer Using Waste Heat Boiler, 1600°F Combustion Temperature, 0.75 sec Residence Time, and Heat Contents from 1 to 50 Btu/scf

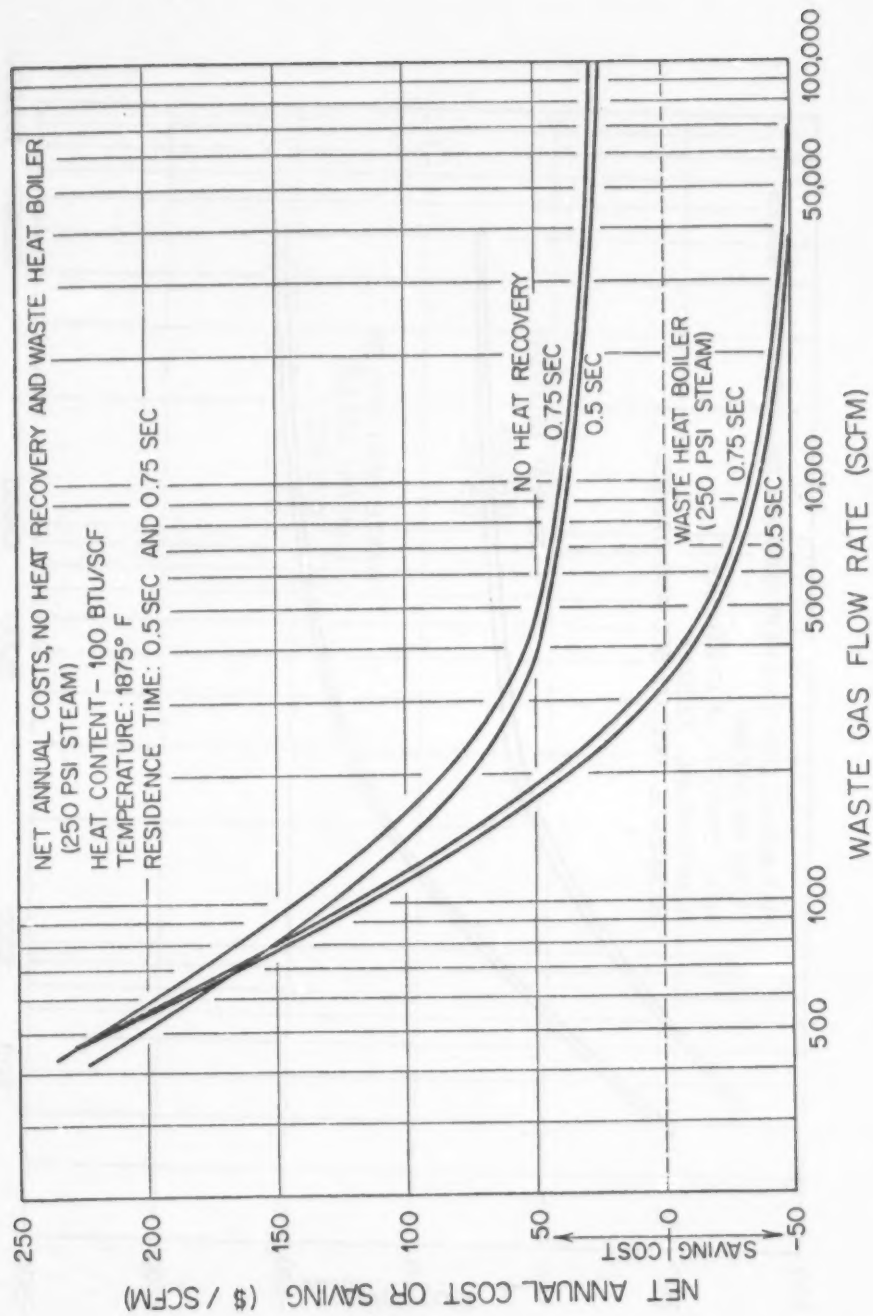


Fig. V-30. Net Annual Cost vs Waste Gas Flow Rate for Thermal Oxidizer with No Heat Recovery and with a Waste Heat Boiler; Heat Content = 100 Btu/scf

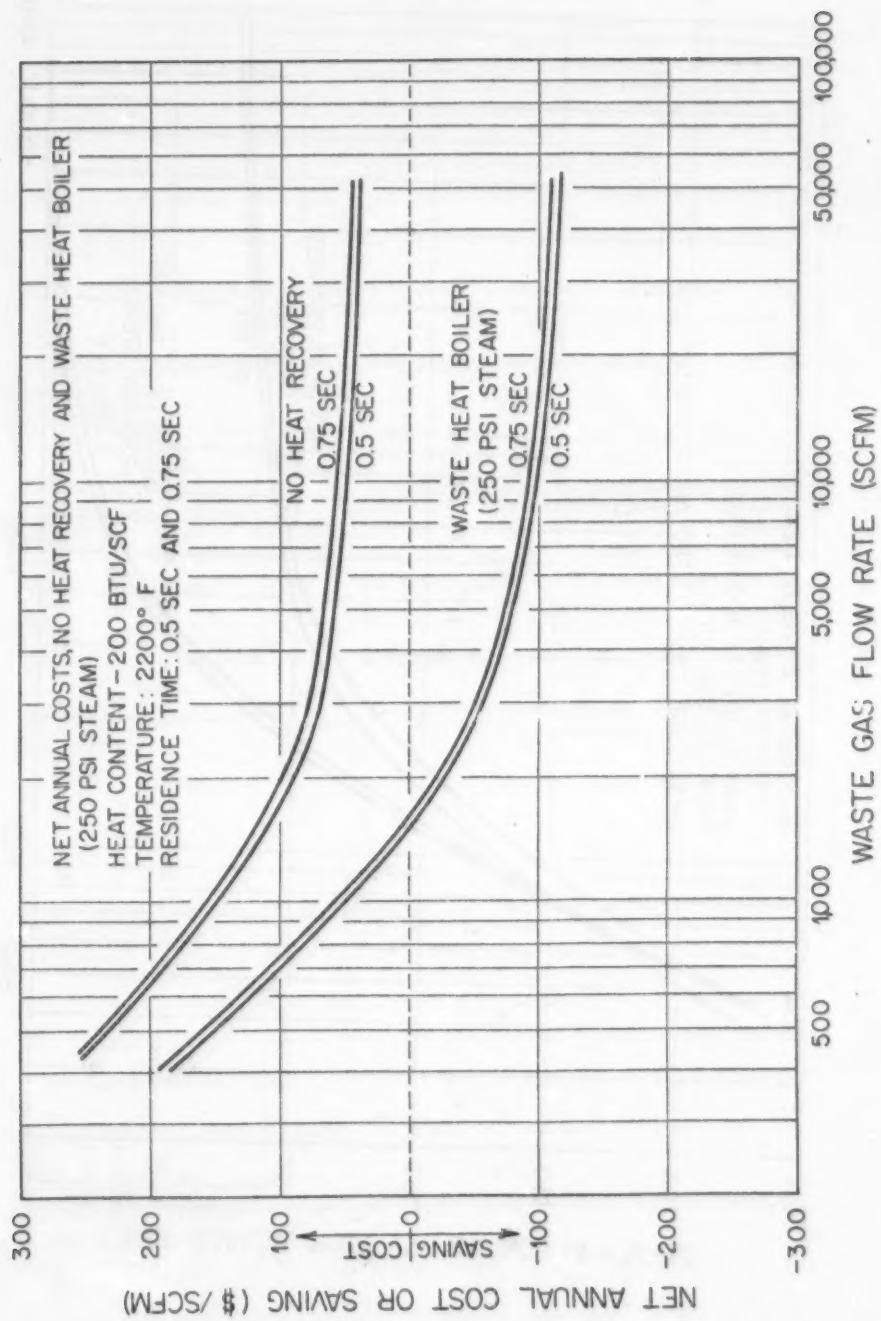


Fig. V-31. Net Annual Cost vs Waste Flow Rate for Thermal Oxidizer with No Heat Recovery and with a Waste Heat Boiler; Heat Content = 200 Btu/scf

Figures V-18 through V-31 present the annual cost of thermal oxidation for various cases.

D. COST EFFECTIVENESS AND ENERGY EFFECTIVENESS

The cost effectiveness and energy effectiveness are calculated by dividing the annual cost for a particular option (Appendix B) or the fuel usage (in Btu/yr) by the total annual amount of VOC destroyed, with the destruction efficiencies assumed as given in Table II-2. Changes in these values with changes in destruction efficiencies owing to specific applications different from the conservative design used in this report are small.

The cost effectiveness is presented in Table V-3 and the energy effectiveness is presented in Table V-4. Cost-effectiveness graphs are shown as a function of destruction efficiency in Figs. V-32 through V-34. Data on cases not shown in the above-mentioned tables and figures can be easily developed by use of Appendix B.

E. OTHER IMPACTS

Other than costs and energy consumption, other impacts of thermal oxidation must be related to the flue gas and the components it emits to the environment. A typical analysis of the flue gas with the emission ratio (lb of component/1000 scf of waste gas) is given in Table V-5.

Table V-3. Cost Effectiveness of Thermal Oxidation

Waste Gas Heat Content (Btu/scf)	Waste Gas Flow Rate (scfm)	Cost Effectiveness (\$ per lb of VOC Destroyed)							
		VOC Destroyed (lb/hr) ^c at 90%	VOC Destroyed (lb/hr) ^c at 99%	90% VOC Destruction ^a				99% VOC Destruction ^b	
				Case I No Heat Recovery	Case II-50 50% Recuperative Heat Recovery	Case III-250 Waste Heat Boiler 250 psi Steam	Case I No Heat Recovery	Case II-50 50% Recuperative Heat Recovery	Case III-250 Waste Heat Boiler 250 psi Steam
1	700	2.59	2.84	\$6.39	\$6.35	\$6.61	\$6.43	\$6.07	\$6.31
	5,000	18.5	20.3	2.83	2.00	2.01	3.19	2.01	2.08
	50,000	184.9	203.3	2.29	1.26	1.25	2.67	1.32	1.35
	100,000	369.7	406.7	2.25	1.20	1.19	2.63	1.28	1.28
10	700	25.9	28.4	0.546	0.534	0.566	0.615	0.575	0.603
	5,000	184.9	203.3	0.253	0.159	0.170	0.291	0.169	0.180
	50,000	1,849	2,033	0.199	0.0849	0.0950	0.239	0.100	0.107
	100,000	3,697	4,066	0.195	0.0794	0.0888	0.235	0.096	0.100
13	700	33.6	37.0	0.459	0.452	0.476	0.463	0.435	0.454
	5,000	240.3	264.3	0.187	0.114	0.123	0.217	0.122	0.132
	50,000	2,403	2,643	0.145	0.0568	0.0651	0.177	0.0693	0.0754
	100,000	4,806	5,287	0.142	0.0527	0.0603	0.174	0.0662	0.0702
20	700	51.8	57.0	0.284	0.284	0.295	0.288	0.266	0.282
	5,000	369.7	406.7	0.109	0.0648	0.0679	0.130	0.0635	0.0747
	50,000	3,697	4,067	0.0821	0.0277	0.0303	0.104	0.0341	0.0381
	100,000	7,394	8,134	0.0801	0.0250	0.0272	0.102	0.0321	0.0347
50	700	129.4	142.4	0.0944	d	0.0988	0.108	d	0.105
	5,000	924.3	1,017	0.0232		0.0067	0.0334		0.0113
	50,000	9,243	10,170	0.0123		(0.0084)	0.0233		(0.0033)
	100,000	18,486	20,340	0.0115		(0.0096)	0.0223		(0.0046)
100	500	e	203.4				0.0572	f	0.0601
	5,000		2,034				0.0123		(0.0057)
	50,000		20,340				0.0069		(0.0143)
	100,000		40,670				0.0065		
200	500	e	406.7				0.0326	f	0.0211
	5,000		4,067				0.0082		(0.0011)
	50,000		40,670				0.0054		(0.0162)

^a Assumes 1400°F combustion temperature and 0.5-sec residence time except where otherwise noted.

^b Assumes 1600°F combustion temperature and 0.5-sec residence time except where otherwise noted.

^c VOC molecular weight, 50; molar heat of combustion, 730,250 Btu/lb-mole of VOC.

^d 50% recuperative heat recovery is not applicable to heat contents of 50 Btu/scf.

^e >1600°F combustion temperature assumes only 99% efficiency case.

^f Recuperative heat recovery is not applicable to temperatures >1600°F.

Table V-4. Fuel Energy Effectiveness of Thermal Oxidation

Waste Gas Heat Content (Btu/scf) ^a	Fuel Energy Usage (Btu/scf)		Net Energy Usage ^b (Btu/scf)		Energy Effectiveness (Btu/lb of VOC Destroyed)			
	99% VOC ^c Destruction		99% VOC ^c Destruction		90% VOC Destruction ^c		99% VOC Destruction ^d	
	90% VOC ^c Destruction	99% VOC ^c Destruction	90% VOC ^c Destruction	99% VOC ^c Destruction	Case I No Heat Recovery	Case III-400 Waste Heat Boiler 400-psi Steam	Case I No Heat Recovery	Case III-400 Waste Heat Boiler 400-psi Steam
1	63.5	82.3	26.3	30.6	927,400	384,100	1,202,000	446,900
10	54.2	73.0	17.0	21.3	79,200	24,800	106,600	31,100
13	51.0	69.9	13.8	18.2	57,300	15,500	78,500	20,400
20	43.6	62.5	6.37	10.8	31,800	4,700	45,600	7,900
50	12.0	31.1	(25.2)	(20.6)	3,500	(7,400)	9,100	(6,000)
100		13.2 ^e	f	(68.4)	f	f	1,900	(10,000)
200		20.3 ^e	f	(163.4)	f	f	1,500	(11,900)

^a scf of waste gas.^b Based on waste heat boiler generating 400-psi steam.^c 1400°F and 0.5-sec residence time.^d 1600°F and 0.5-sec residence time.^e 1600°F and 0.5-sec residence time.^f 99% VOC destruction only at >1600°F and 0.5-sec residence time.

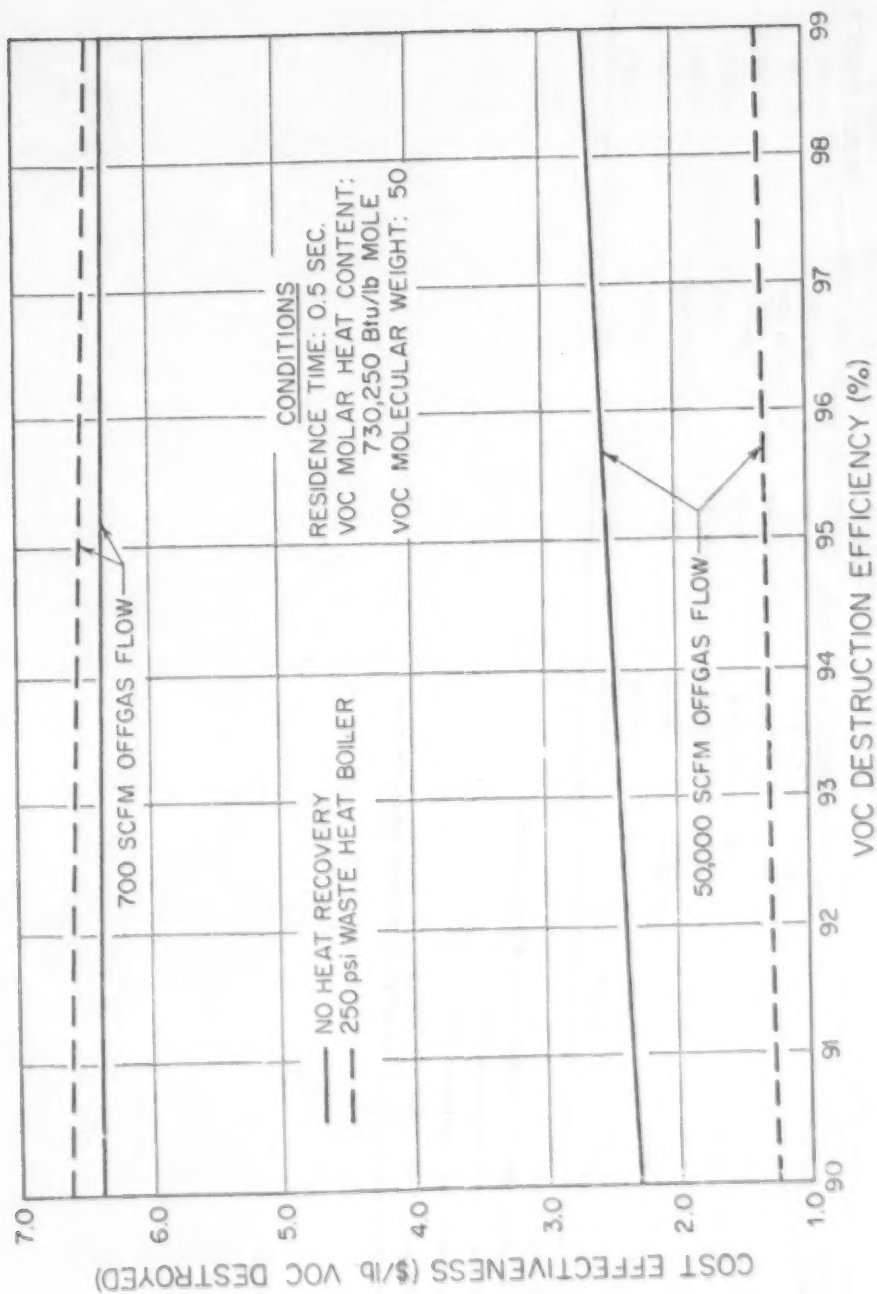


Fig. V-32. Cost Effectiveness vs VOC Destruction Efficiency for Waste Gases with 1 Btu/scf Heat Content

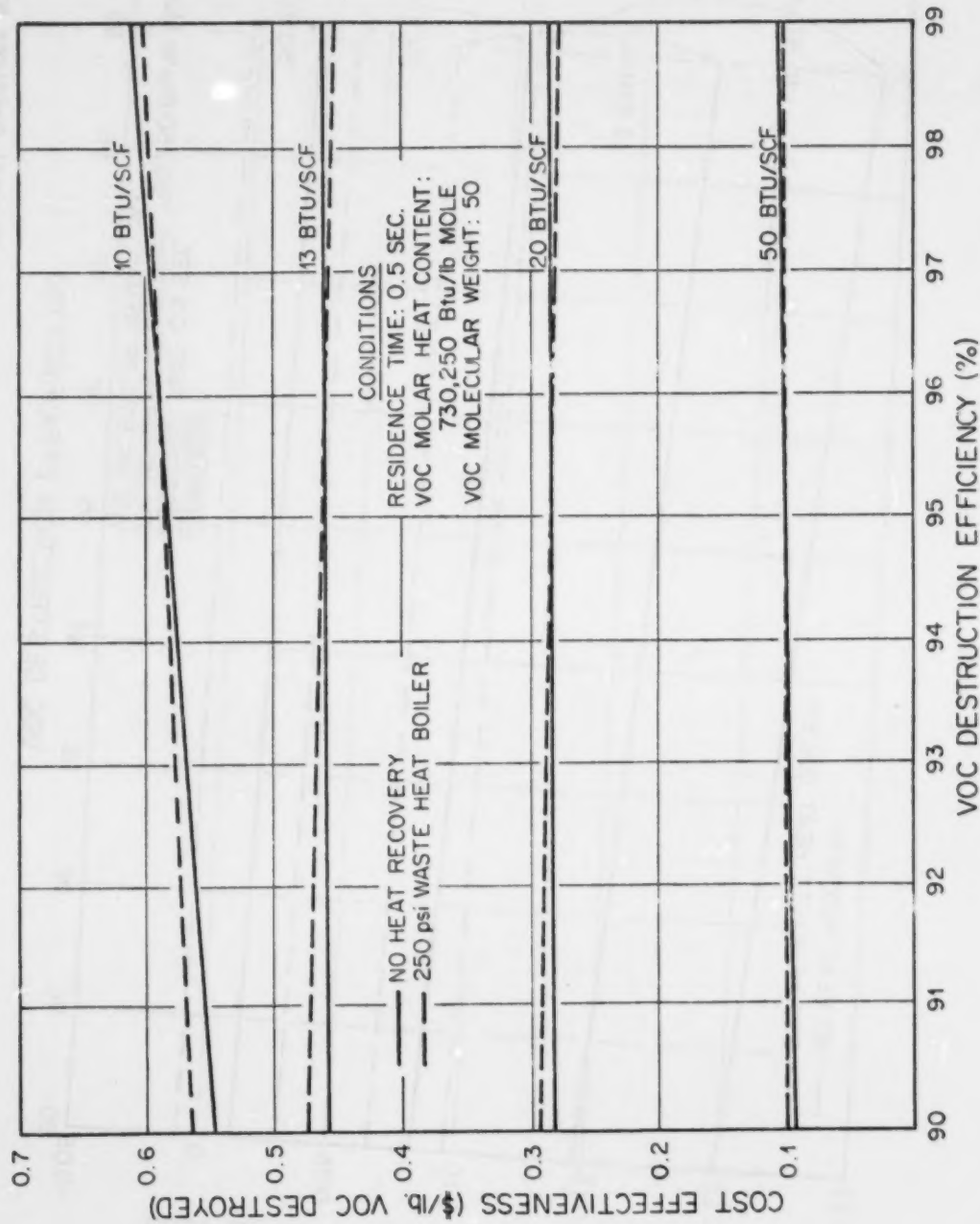


Fig. V-33. Cost Effectiveness vs VOC Destruction Efficiency for Thermal Oxidizer with No Heat Recovery and with a Waste Heat Boiler; Heat Contents = Between 1 and 50 Btu/scf and Waste Gas Flows = 700 scfm

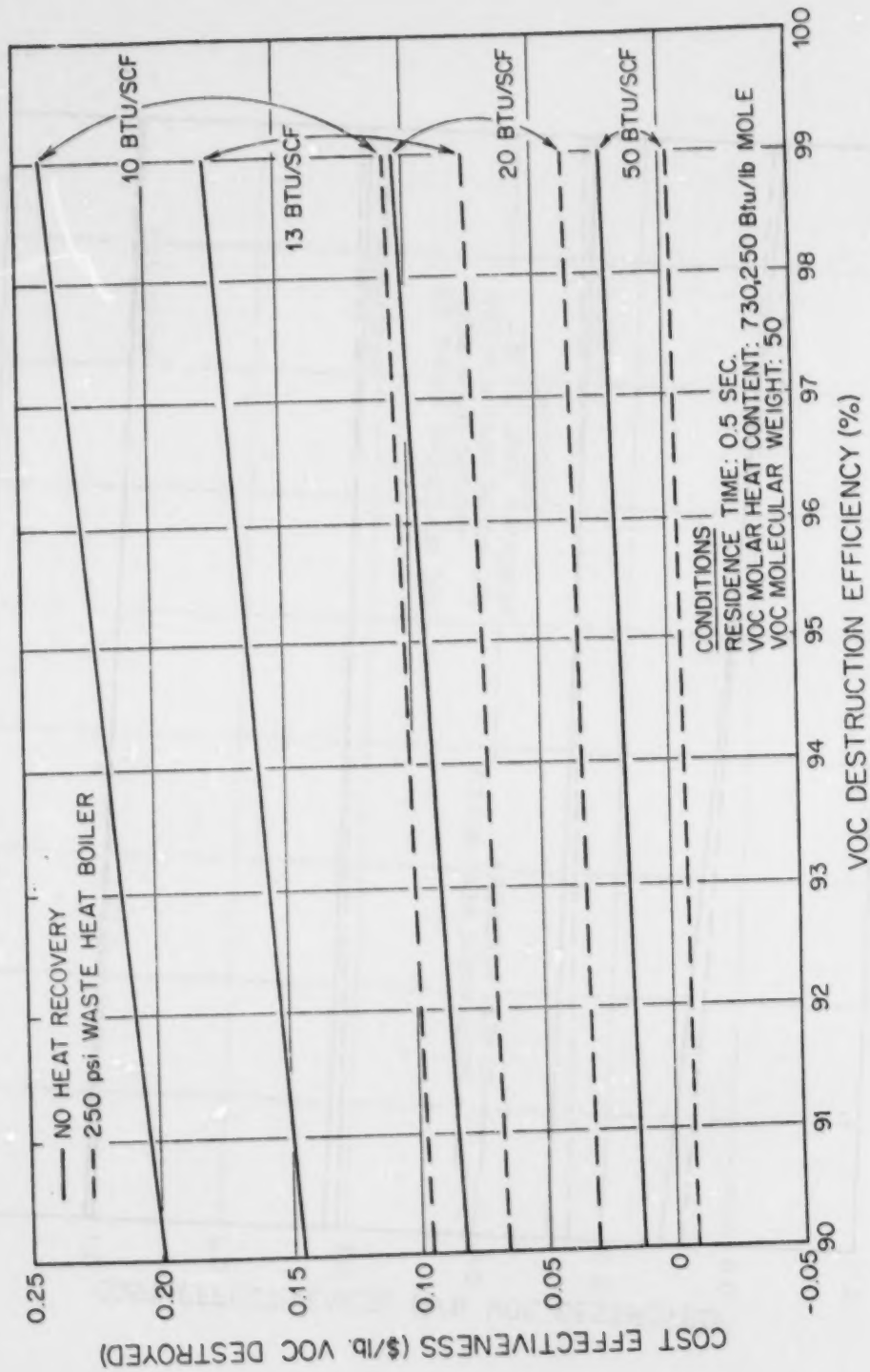


Fig. V-34. Cost Effectiveness vs VOC Destruction Efficiency for Thermal Oxidizer with No Heat Recovery and with a Waste Heat Boiler; Heat Contents = Between 1 and 50 Btu/scf and Waste Gas Flows = 50,000 scfm

Table V-5. Typical Thermal Oxidizer Flue Gas Composition

Component	Composition	Emission Ratio (lb/1000 scf of Waste Gas)
CO ₂ ^a	4.0 vol %	2.2
H ₂ O ^a	10.0 vol %	12.2
N ₂ ^a	82.9 vol %	65.1
O ₂ ^a	3.0 vol %	2.1
VOC and CH ₄ ^b	500 - 50 ppm	0.0022 - 0.0002
Carbon monoxide ^c	10 - 6 ppm	0.001 - 0.0005
Particulates ^c (as carbon)	30 - 3 ppm	0.001 - 0.0001
Sulfur oxides ^c (as SO ₂)	<1 ppm	0.00005 - 0.00001
Nitrogen oxides (as NO ₂) ^c	160 - 30 ppm	0.02 - 0.004

^a Calculated from thermal oxidizer heat and material balance for a low- to moderate-heat-content waste gas.

^b Calculated for 90 and 99% total VOC destruction efficiency.

^c Calculated from natural combustion emission factors for industrial boilers reported in AP-42, Supplement 7 (ref. 12). Variations relate to variations in factors reported and/or variations in fuel usage for low- to moderate-heat-content waste gases.

Carbon monoxide, particulates, sulfur oxides, and nitrogen oxide emissions are small if these agents form in similar quantities to that reported for industrial gas boilers.¹² This assumption may be adequate since the fuel burner region approximates the fuel burners in boilers. This, however, has not been demonstrated with experimental data. As cautioned in AP-42, these values are sensitive to burner upsets and may increase by several orders of magnitude in a poorly designed or operated unit. Increases in these compounds may be caused by waste gases containing high levels of halogens, sulfur, or nitrogen containing VOC compounds.

This study does not quantitatively assess the relationship between the VOC destruction efficiency and the impacts of the components emitted after thermal oxidation. However, qualitative trends do exist. For instance, a higher VOC destruction efficiency means a high combustion temperature (or a longer residence time). A higher combustion temperature may increase nitrogen oxides to an unacceptable level. However, increased temperatures with adequate oxygen should decrease any particulate formation if it is carbonaceous in nature.

VI. SUMMARY AND CONCLUSIONS

Thermal oxidation is a widely used control technique for VOC emissions. The limits and design principles of this technique are evaluated in this report.

Design criterion and design procedures are presented that allow for a preliminary thermal oxidizer design. Thermal oxidizers without heat recovery and with five levels of recuperative and waste-heat steam boiler heat recovery are considered. Capital and operating costs are developed, and the annual cost of thermal oxidation is calculated as functions of the characteristics of the waste gas. Cost effectiveness and energy effectiveness of 90 and 99% VOC destruction efficiencies are developed.

The conclusions derived from the cost evaluation are as follows:

1. Since the thermal oxidizer design used here is quite conservative, the cost-related parameters actually experienced in industry are expected to impose a lesser economic hardship than is presented.
2. The waste gas heat content is a highly sensitive variable in determining annual costs, cost effectiveness, and energy effectiveness. In general, as the heat content increases, the annual costs, cost effectiveness, and the energy effectiveness decrease. This leads to the general statement that waste gases with higher heat contents (same flow) cost less to control than those with lower heat contents.
3. The waste gas flow rate is a highly sensitive variable in determining annual costs and cost effectiveness. Energy effectiveness is independent of the flow rate. As the waste gas flow increases (at a constant heat content), the annual costs increase but the annual cost per scfm of waste gas, the cost/feed flow ratio, decreases. This ratio decreases drastically between low flows (700 scfm) and moderate flows (5000 scfm), but remains relatively constant between moderate (5000 scfm) to large flows (50 to 100,000 scfm). The increase of waste gas flow favorably decreases the cost effectiveness of control much like the annual cost/feed flow ratio noted above. Energy effectiveness is constant with flow.
4. Annual net costs of control decrease as the level of heat recovery increases for the same waste gas.

5. The annual cost of systems using no heat recovery is normally higher than that for systems using maximum heat recovery. This may not apply to waste gases with low flow rates (700 scfm).
6. The differences in annual costs (all heat contents and flows) between recuperative heat recovery at 70% and waste heat boilers generating steam at 250 and 400 psi are small. The waste heat boiler case (250 psi) may be considered to have the lowest possible annual cost for a given heat constant and flow rate.
7. Residence time increases of from 0.5 to 0.75 sec have little effect on the annual costs or cost effectiveness, and no effect on the energy effectiveness. Lower combustion temperatures, however, reduce the quantity of energy recoverable from the system.
8. Increasing the VOC destruction efficiency from 90 to 99% by raising the combustion temperature increases the cost effectiveness and energy effectiveness for all cases except the cost effectiveness of low flow waste gas with maximum heat recovery used. Social costs are not considered.
9. Increasing the VOC destruction efficiency through increased residence time may be more cost effective and energy effective than increasing the efficiencies by increasing the combustion temperatures.

VII. REFERENCES*

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*When a reference number is used at the end of a paragraph or on a heading, it usually refers to the entire paragraph or material under the heading. When, however, an additional reference is required for only a certain portion of the paragraph or captioned material, the earlier reference number may not apply to that particular portion.

APPENDIX A

PURCHASE COSTS FOR THERMAL OXIDATION COMBUSTION
CHAMBERS, RECUPERATIVE HEAT EXCHANGERS,
AND WASTE HEAT BOILERS

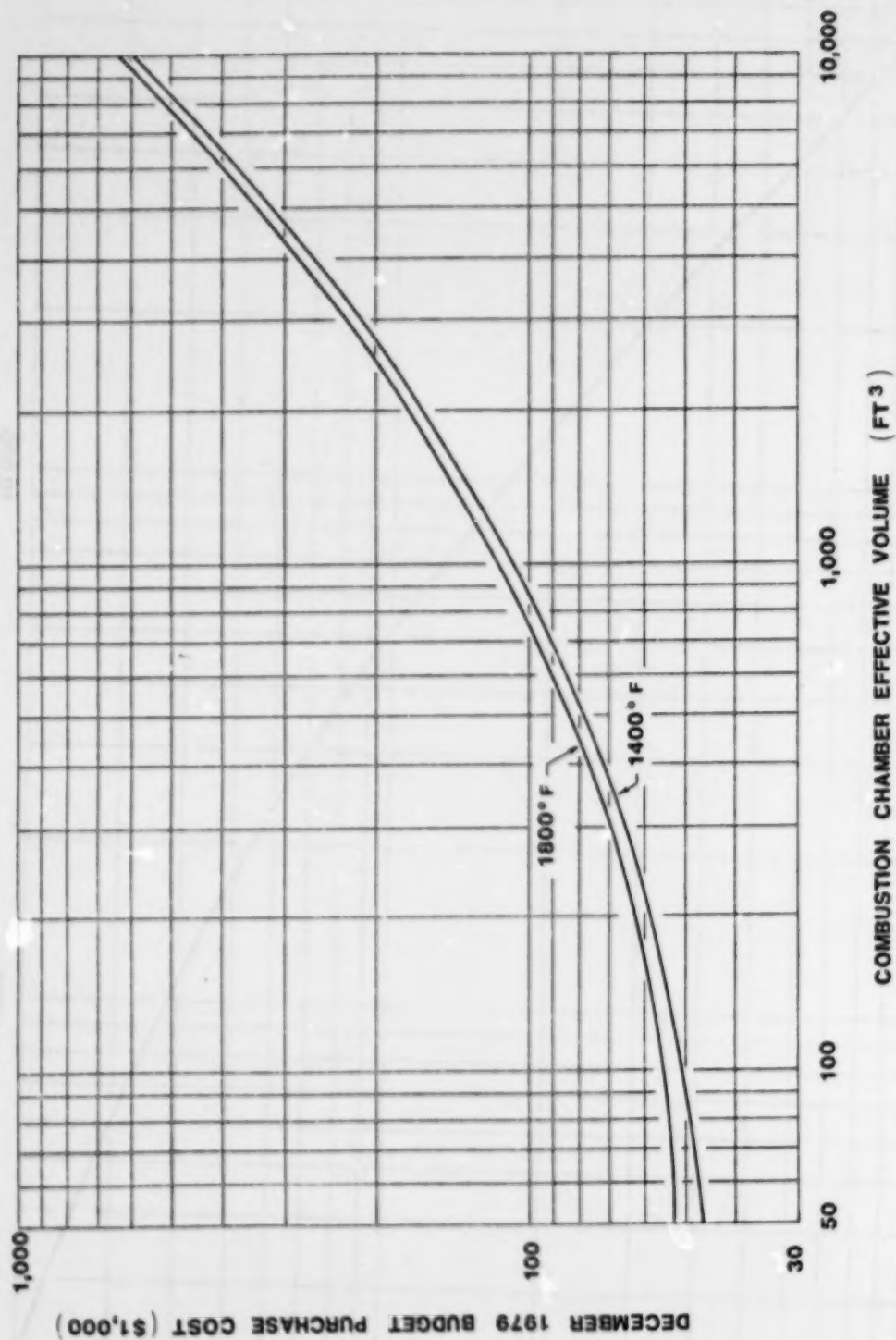


Fig. A-1. Purchase Costs for Thermal Oxidation Combustion Chambers

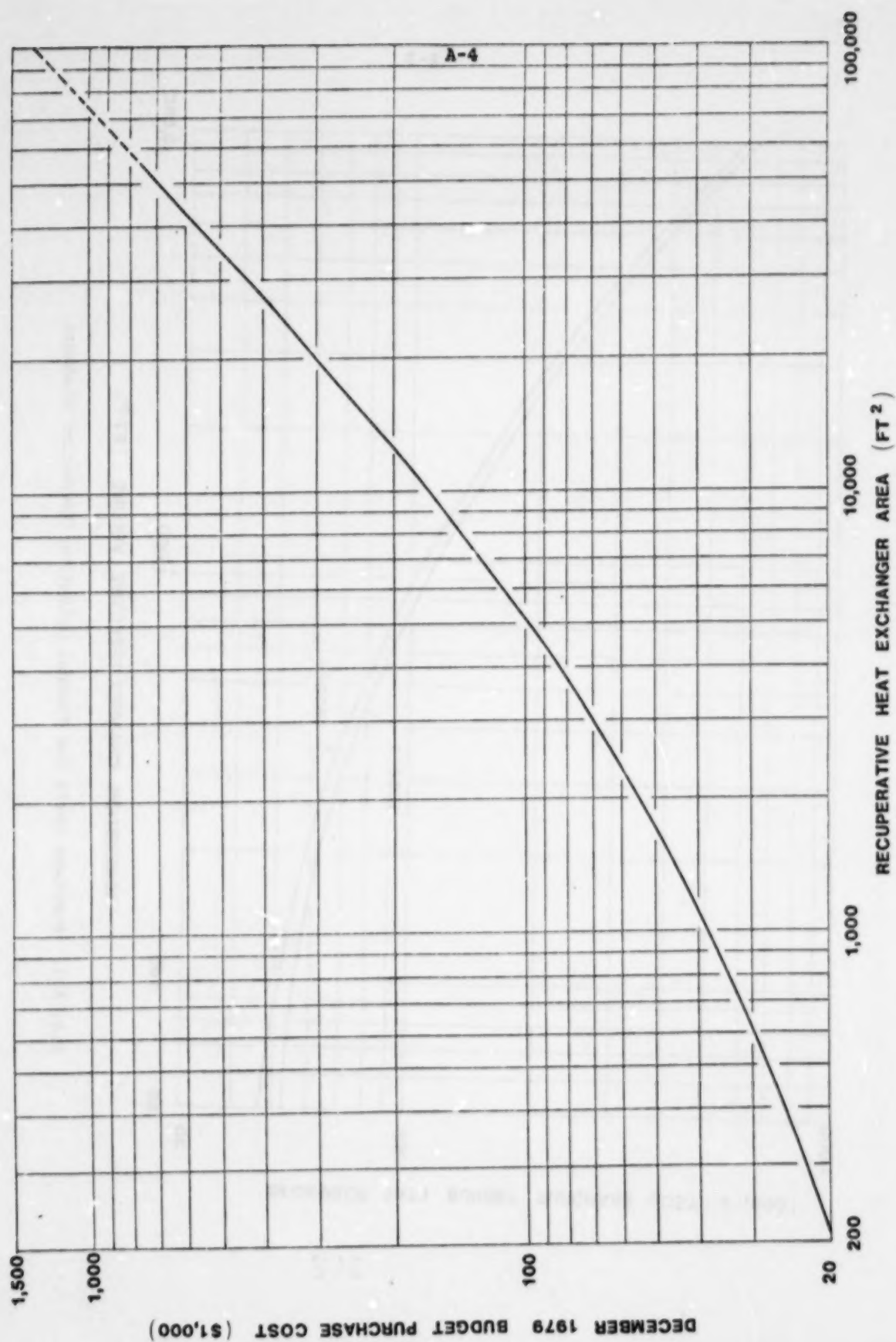


Fig. A-2. Purchase Costs for Thermal Oxidation Recuperative Heat Exchangers

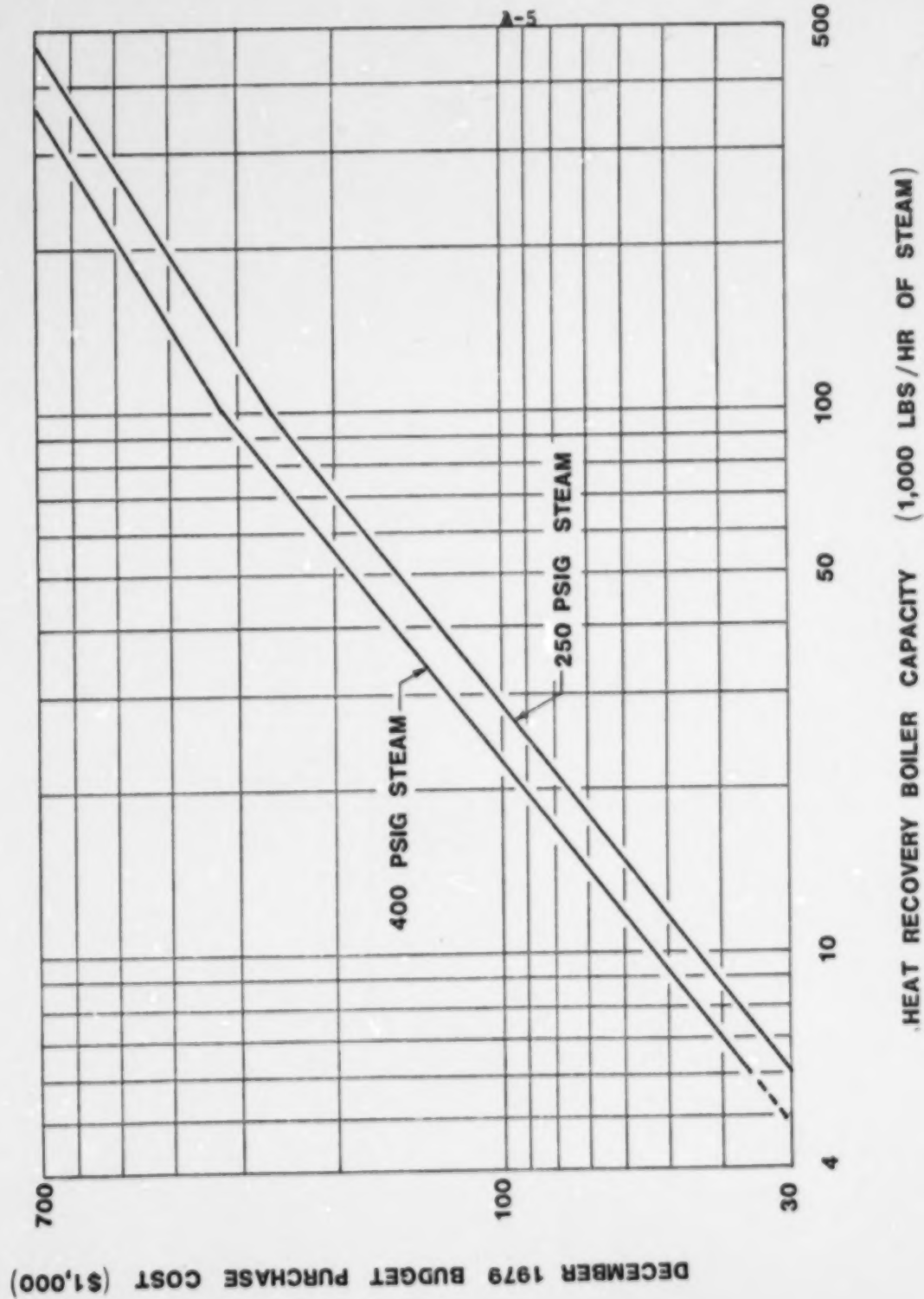


Fig. A-3. Purchase Costs for Thermal Oxidation Waste Heat Boilers

APPENDIX B

ANNUAL COST DATA

OFF GAS HEAT CONTENT 1 BTU/SCF.
 RESIDENCE TIME 0.5 SEC.
 COMB. TEMP 1400 °F

CASE	OFF GAS FLOW SCFM	CAPITAL COST \$1000	OPERATING COSTS-OR- CREDIT			NET ANNUALIZED COST - OR - SAVINGS \$1000	NET COST OR SAVING PER SCFM \$/SCFM
			FIXED COSTS \$1000	UTILITIES \$1000	MANPOWER \$1000		
I NO HEAT RECOVERY	700	272	79	48	18	145	207.1
	5000	359	104	337	18	459	91.8
	50000	1105	320	3367	18	3705	74.1
	100000	1810	525	6734	18	7277	72.8
II-30 30% HEAT RECOVERY <i>reheat</i>	700	313	91	30	22	143	204.3
	5000	472	137	214	22	373	74.6
	50000	1480	429	2147	22	2598	52.0
	100000	2493	723	4295	22	5040	50.4
II-50 50% HEAT RECOVERY <i>reheat</i>	700	333	97	20	27	144	205.7
	5000	527	163	144	27	324	64.8
	50000	1960	568	1442	27	2037	40.7
	100000	3400	986	2885	27	3898	39.0
II-70 70% HEAT RECOVERY <i>reheat</i>	700	337	98	12	32	142	202.9
	5000	613	178	85	32	295	59.0
	50000	2800	812	848	32	1692	33.8
	100000	5133	1489	1696	32	3217	32.2
III-400 WASTE HT. BOILER 400PSI STM.	700	339	98	48	36	154	220.0
	5000	538	156	339	"	335	62.0
	50000	2124	616	3386	"	2080	41.9
	100000	3802	1103	6773	"	3996	40.0
III-250 WASTE HT. BOILER 250PSI STM.	700	325	94	48	"	150	214.3
	5000	504	146	339	"	325	65.0
	50000	1940	563	3386	"	2027	40.5
	100000	3308	959	6773	"	3853	38.5

OFF GAS HEAT CONTENT 1 BTU/SCF
 RESIDENCE TIME 0.75 SEC.
 COMB. TEMP 1400 °F

CASE	OFF GAS FLOW SCFM	CAPITAL COST \$1000	OPERATING COSTS-OR-CREDIT			NET ANNUALIZED COST-OR- SAVINGS \$1000	NET COST OR SAVING PER SCFM \$/SCFM	%
			FIXED COSTS \$1000	UTILITIES \$1000	MANPOWER \$1000	RECOVERY CREDIT \$1000		
I NO HEAT RECOVERY	700	273	79	48	18	—	207.1	C
	5000	394	114	337	"	—	93.9	C
	50000	1422	412	3367	"	—	76.0	C
	100000	2376	689	6734	"	—	74.4	C
II-30 30% HEAT RECOVERY <i>reheat</i>	700	314	91	30	22	—	204.3	C
	5000	495	144	214	"	—	76.0	C
	50000	1733	503	2147	"	—	53.4	C
	100000	3000	870	4295	"	—	51.9	C
II-50 50% HEAT RECOVERY <i>reheat</i>	700	340	99	20	27	—	208.6	C
	5000	545	158	144	"	—	65.8	C
	50000	2173	630	1442	"	—	42.0	C
	100000	3907	1133	2885	"	—	40.5	C
II-70 70% HEAT RECOVERY <i>reheat</i>	700	341	99	12	23	—	204.3	C
	5000	647	188	85	"	—	61.0	C
	50000	3013	874	848	"	—	36.1	C
	100000	5600	1624	1696	"	—	33.5	C
III-400 WASTE HT. BOILER 400PSI STM.	700	339	98	48	26	28	220.0	C
	5000	574	166	339	"	196	69.0	C
	50000	2441	708	3386	"	1458	43.4	C
	100000	4370	1267	6773	"	3915	41.6	C
III-250 WASTE HT. BOILER 250PSI STM.	700	325	94	48	"	28	214.3	C
	5000	541	157	339	"	146	67.2	C
	50000	2257	655	3386	"	1458	42.4	C
	100000	3869	1122	6773	"	3915	40.2	C

OFF GAS HEAT CONTENT 10 BTU/SCF.
 RESIDENCE TIME 0.5 SEC.
 COMB. TEMP 1400 °F

CASE	OFF GAS FLOW SCFM	CAPITAL COST \$1000	OPERATING COSTS-OR-CREDIT			NET ANNUALIZED COST - OR - SAVINGS \$1000	NET COST OR SAVING PER SCFM \$/SCFM	C
			FIXED COSTS \$1000	UTILITIES \$1000	MANPOWER \$1000	RECOVERY CREDIT \$1000		
I NO HEAT RECOVERY	700	272	79	41	18	—	138	C
	5000	359	104	288	"	—	410	C
	50000	1105	320	2878	"	—	3216	C
	100000	1810	525	5756	"	—	6299	C
II-30 30% HEAT RECOVERY	700	313	91	22	22	—	135	C
	5000	472	137	154	"	—	313	C
	50000	1480	429	1528	"	—	1979	C
	100000	2493	723	3055	"	—	3800	C
II-50 50% HEAT RECOVERY	700	333	97	11	27	—	135	C
	5000	527	163	78	"	—	258	C
	50000	1960	568	780	"	—	1375	C
	100000	3400	986	1560	"	—	2573	C
II-70 70% HEAT RECOVERY	700	337	98	5	32	—	135	C
	5000	613	178	31	"	—	241	C
	50000	2800	812	2612	"	—	1166	C
	100000	5133	1489	624	"	—	2145	C
III-400 WASTE HT. BOILER 400PSI, STM.	700	339	98	41	36	28	147	C
	5000	538	156	290	"	196	286	C
	50000	2124	616	2897	"	1958	1591	C
	100000	3802	1103	5795	"	3915	3019	C
III-250 WASTE HT. BOILER 250PSI, STM.	700	325	94	41	"	28	143	C
	5000	504	146	290	"	196	276	C
	50000	1940	563	2897	"	1958	1538	C
	100000	3308	959	5795	"	3915	2875	C

OFF GAS HEAT CONTENT 10 BTU/SCF.
 RESIDENCE TIME 0.75 SEC.
 COMB. TEMP 1400 °F

CASE	OFF GAS FLOW SCFM	CAPITAL COST \$1000	OPERATING COSTS-OR-CREDIT			NET ANNUALIZED COST-OR- SAVINGS \$1000	NET COST OR SAVING PER SCFM	
			FIXED COSTS \$1000	UTILITIES \$1000	MANPOWER \$1000		\$/SCFM	%
I NO HEAT RECOVERY	700	273	79	41	18	—	197.1	C
	5000	394	114	288	"	—	84.0	C
	50000	1420	412	2878	"	—	66.2	C
	100000	2376	689	5756	"	—	64.6	C
II-30 30% HEAT RECOVERY <i>assumed</i>	700	314	91	22	22	—	192.9	C
	5000	495	144	154	"	—	64.0	C
	50000	1733	503	1528	"	—	41.1	C
	100000	3000	870	3055	"	—	39.5	C
II-50 50% HEAT RECOVERY <i>assumed</i>	700	340	99	11	27	—	195.7	C
	5000	545	158	78	"	—	52.6	C
	50000	2173	630	780	"	—	28.7	C
	100000	3907	1133	1560	"	—	27.2	C
II-70 70% HEAT RECOVERY <i>assumed</i>	700	341	99	5	32	—	194.3	C
	5000	647	188	31	"	—	50.2	C
	50000	3013	874	312	"	—	24.4	C
	100000	5600	1624	624	"	—	22.8	C
III-400 WASTE HT. BOILER 400 PSI. STM.	700	339	98	41	36	28	210.0	C
	5000	574	166	290	"	196	59.2	C
	50000	2441	708	2897	"	1968	33.7	C
	100000	4370	1267	5795	"	3915	31.8	C
III-250 WASTE HT. BOILER 250 PSI. STM.	700	326	95	41	"	28	205.7	C
	5000	541	157	290	"	196	57.4	C
	50000	2257	655	2897	"	1958	32.6	C
	100000	3869	1122	5795	"	3915	30.4	C

OFF GAS HEAT CONTENT 13 BTU/SCF.
 RESIDENCE TIME 0.5 SEC.
 COMB. TEMP 1400 °F

CASE	OFF GAS FLOW SCFM	CAPITAL COST \$1000	OPERATING COSTS-OR-CREDIT				NET ANNUALIZED COST-OR- SAVINGS \$1000	NET COST OR SAVING PER SCFM	
			FIXED COSTS \$1000	UTILITIES \$1000	MANPOWER \$1000	RECOVERY CREDIT \$1000		\$/SCFM	%
I NO HEAT RECOVERY	700	272	79	38	18	—	135	192.9	C
	5000	359	104	271	"	—	393	78.6	C
	50000	1105	320	2710	"	—	3048	61.0	C
	100000	1810	525	5420	"	—	5963	59.6	C
II-30 30% HEAT RECOVERY	700	313	91	19	22	—	132	188.6	C
	5000	472	137	134	"	—	293	58.6	C
	50000	1480	429	1343	"	—	1794	35.9	C
	100000	2493	723	2687	"	—	3432	34.3	C
II-50 50% HEAT RECOVERY	700	333	97	9	27	—	133	190.0	C
	5000	527	153	60	"	—	240	48.0	C
	50000	1960	568	601	"	—	1196	23.9	C
	100000	3400	986	1204	"	—	2217	22.2	C
II-70 70% HEAT RECOVERY	700	337	98	5	32	—	135	192.9	C
	5000	613	178	31	"	—	241	48.2	C
	50000	2800	812	312	"	—	1156	23.1	C
	100000	5133	1489	624	"	—	2145	21.5	C
III-400 WASTE HT. BOILER 400PSI. STM.	700	339	98	38	36	28	144	205.7	C
	5000	538	156	273	"	196	269	53.8	C
	50000	2124	616	2729	"	1968	1423	28.6	C
	100000	3802	1103	5459	"	3915	2683	26.8	C
III-250 WASTE HT. BOILER 250PSI. STM.	700	325	94	38	"	28	140	200.0	C
	5000	504	146	273	"	196	259	51.8	C
	50000	1940	563	2729	"	1968	1370	27.4	C
	100000	3308	959	5459	"	3915	2539	25.4	C

OFF GAS HEAT CONTENT 13 BTU/SCF.
 RESIDENCE TIME 0.75 SEC.
 COMB. TEMP 1400 °F

CASE	OFF GAS FLOW SCFM	CAPITAL COST \$1000	OPERATING COSTS-OR- CREDIT				NET ANNUALIZED COST-OR- SAVINGS \$1000	NET COST OR SAVING PER SCFM \$/SCFM	%
			FIXED COSTS \$1000	UTILITIES \$1000	MANPOWER \$1000	RECOVERY CREDIT \$1000			
I NO HEAT RECOVERY	700	273	79	38	18	-	135	192.9	C
	5000	394	114	271	"	-	403	80.6	C
	50000	1422	412	2710	"	-	3140	62.8	C
	100000	2376	689	5420	"	-	6127	61.3	C
II-30 30% HEAT RECOVERY	700	314	91	19	22	-	132	188.6	C
	5000	495	144	134	"	-	300	60.0	C
	50000	1733	503	1343	"	-	1868	37.4	C
	100000	3000	870	2687	"	-	3579	35.8	C
II-50 50% HEAT RECOVERY	700	340	99	9	27	-	135	192.9	C
	5000	545	158	60	"	-	245	49.0	C
	50000	2173	630	601	"	-	1258	25.2	C
	100000	3907	1133	1204	"	-	2364	23.6	C
II-70 70% HEAT RECOVERY	700	341	99	5	32	-	136	194.3	C
	5000	647	188	31	"	-	251	50.2	C
	50000	3013	874	312	"	-	1218	24.4	C
	100000	5600	1624	624	"	-	2280	22.8	C
III-400 WASTE HT. BOILER 900PSI 51A.	700	339	98	38	36	28	144	205.7	C
	5000	574	166	273	"	146	279	55.8	C
	50000	2441	708	2729	"	1458	1515	30.3	C
	100000	4370	1267	5459	"	3415	2847	28.5	C
III-250 WASTE HT. BOILER 250PSI 51A.	700	325	94	38	"	28	140	200.0	C
	5000	541	157	223	"	146	270	54.0	C
	50000	2257	655	2729	"	1458	1462	29.2	C
	100000	3869	1122	5459	"	3415	2702	27	C

OFF GAS HEAT CONTENT 20 BTU/SCF
 RESIDENCE TIME 0.5 SEC.
 COMB. TEMP 1400 °F

CASE	OFF GAS FLOW SCFM	CAPITAL COST \$1000	OPERATING COSTS-OR-CREDIT				NET ANNUALIZED COST-OR- SAVINGS \$1000 1%/s	NET COST OR SAVING PER SCFM	
			FIXED COSTS \$1000	UTILITIES \$1000	MANPOWER \$1000	RECOVERY CREDIT \$1000		\$/SCFM	1%/s
I NO HEAT RECOVERY	700	272	79	32	18	—	129	184.3	C
	5000	359	104	232	"	—	354	70.8	C
	50000	1105	320	2321	"	—	2659	63.2	C
	100000	1810	525	4642	"	—	5185	51.9	C
II-30 30% HEAT RECOVERY	700	313	91	14	22	—	127	181.4	C
	5000	472	137	94	"	—	253	50.6	C
	50000	1480	429	945	"	—	1396	27.9	C
	100000	2443	723	1888	"	—	2633	26.3	C
II-50 50% HEAT RECOVERY	700	333	97	5	27	—	129	184.3	C
	5000	627	153	30	"	—	210	42.0	C
	50000	1960	568	302	"	—	897	17.4	C
	100000	3400	986	605	"	—	1618	16.2	C
II-70 70% HEAT RECOVERY	700								
	5000								
	50000								
	100000								
III-400 WASTE HT. BOILER 400PSI STM.	700	339	98	32	36	28	138	197.1	C
	5000	538	156	234	"	196	230	46.0	C
	50000	2124	616	2340	"	1958	1034	20.7	C
	100000	3802	1103	4681	"	3915	1905	19.0	C
III-250 WASTE HT. BOILER 250PSI STM.	700	325	94	32	"	28	134	191.4	C
	5000	504	146	234	"	196	220	44.0	C
	50000	1910	563	2340	"	1958	981	19.6	C
	100000	3308	959	4681	"	3915	1761	17.6	C

OFF GAS HEAT CONTENT 20 BTU/SCF.
 RESIDENCE TIME 0.75 SEC.
 COMB. TEMP 1400 °F

CASE	OFF GAS FLOW SCFM	CAPITAL COST \$1000	OPERATING COSTS-OR- CREDIT				NET ANNUALIZED COST - OR - SAVINGS \$1000	NET COST OR SAVING PER SCFM \$/SCFM	%
			FIXED COSTS \$1000	UTILITIES \$1000	MANPOWER \$1000	RECOVERY CREDIT \$1000			
I NO HEAT RECOVERY	700	273	79	32	18	—	129	184.3	C
	5000	394	114	232	"	—	364	72.9	C
	50000	1422	412	2321	"	—	2751	55.0	C
	100000	2376	689	4642	"	—	5349	53.5	C
II-30 30% HEAT RECOVERY <i>not paid</i>	700	314	91	14	22	—	127	181.4	C
	5000	495	144	94	"	—	260	52.0	C
	50000	1733	503	945	"	—	1470	29.4	C
	100000	3000	870	1888	"	—	2780	27.8	C
II-50 50 % HEAT RECOVERY <i>not paid</i>	700	340	99	5	27	—	131	187.1	C
	5000	545	158	30	"	—	215	43.0	C
	50000	2173	630	302	"	—	959	19.2	C
	100000	3907	1133	605	"	—	1765	17.7	C
II-70 70% HEAT RECOVERY <i>not paid</i>	700								
	5000								
	50000								
	100000								
III-400 WASTE HT. BOILER 400PSI STIA.	700	339	98	32	36	28	138	197.1	C
	5000	574	166	234	"	196	240	48.0	C
	50000	2441	708	2340	"	1958	1126	22.5	C
	100000	4370	1267	4681	"	3915	2069	20.7	C
III-250 WASTE HT. BOILER 250PSI STIA.	700	325	94	32	"	28	134	191.4	C
	5000	541	157	234	"	196	231	46.2	C
	50000	2267	655	2340	"	1958	1073	21.5	C
	100000	3869	1122	4681	"	3915	1924	19.2	C

OFF GAS HEAT CONTENT 50 BTU/SCF
 RESIDENCE TIME 0.5 SEC.
 COMB. TEMP 1400 °F

CASE	OFF GAS FLOW SCFM	CAPITAL COST \$1000	OPERATING COSTS-OR-CREDIT				NET ANNUALIZED COST-OR-SAVINGS \$1000	NET COST OR SAVING PER SCFM \$/SCFM	C
			FIXED COSTS \$1000	UTILITIES \$1000	MANPOWER \$1000	RECOVERY CREDIT \$1000			
I NO HEAT RECOVERY	700	272	79	10	18	—	107	152.9	C
	5000	359	104	66	"	—	188	37.6	C
	50000	1105	320	660	"	—	998	20.0	C
	100,000	1810	525	1320	"	—	1863	18.6	C
II-30 30% HEAT RECOVERY	700								
	5000								
	50000								
	100,000								
II-50 50% HEAT RECOVERY	700								
	5000								
	50000								
	100,000								
II-70 70% HEAT RECOVERY	700								
	5000								
	50000								
	100,000								
III-400 WASTE HT. BOILER 400PSI STM.	700	339	98	10	36	28	116	165.7	C
	5000	538	156	68	"	196	64	12.8	C
	50000	2124	616	679	"	1958	627	12.5	S
	100,000	3802	1103	1359	"	3915	1417	14.7	S
III-250 WASTE HT. BOILER 250PSI STM.	700	325	94	10	"	28	112	160.0	C
	5000	504	146	68	"	196	64	10.8	C
	50000	1940	563	678	"	1958	680	13.6	S
	100,000	3308	959	1359	"	3915	1661	15.6	S

OFF GAS HEAT CONTENT 50 BTU/SCF
 RESIDENCE TIME 0.75 hr.
 COMB. TEMP 1400 °F

CASE	OFF GAS FLOW SCFM	CAPITAL COST \$1000	OPERATING COSTS-OR-CREDIT				NET ANNUALIZED COST - OR - SAVINGS \$1000	NET COST OR SAVING PER SCFM \$/SCFM	10/5
			FIXED COSTS \$1000	UTILITIES \$1000	MANPOWER \$1000	RECOVERY CREDIT \$1000			
I NO HEAT RECOVERY	700	273	79	10	18	—	107	162.9	C
	5000	394	114	66	"	—	198	39.6	C
	50000	1422	412	660	"	—	1090	21.8	C
	100000	2376	689	1320	"	—	2027	20.3	C
II-30 30% HEAT RECOVERY	700								
	5000								
	50000								
	100000								
II-50 50% HEAT RECOVERY	700								
	5000								
	50000								
	100000								
II-70 70% HEAT RECOVERY	700								
	5000								
	50000								
	100000								
III-400 WASTE HT. BOILER 400 PSI STMA	700	339	98	10	36	28	116	165.7	C
	5000	574	166	68	"	196	74	14.8	C
	50000	2441	708	679	"	1958	535	10.7	S
	100000	4370	1267	1359	"	3915	1283	12.8	S
III-250 WASTE HT. BOILER 250 PSI STMA	700	326	95	10	"	28	113	161.4	C
	5000	541	157	68	"	196	65	13.0	C
	50000	2267	655	679	"	1958	588	11.8	S
	100000	3869	1122	1359	"	3915	1398	14.0	S

1000 BTU/GAS HEAT CONTENT 1 BTU/SCF
 100 SOURCE TIME 0.5 SEC.
 10000 TEMP 1600 °F

GAS FLOW SCFM	CAPITAL COST \$1000	OPERATING COSTS-OR-CREDIT				INITIALIZED COST-OR- SAVINGS \$1000	INITIALIZED SAVINGS \$1000	NET COST-OR- SAVING PER SCFM
		FIXED COSTS \$1000	UTILITIES \$1000	MANPOWER \$1000	RECOVERY CREDIT \$1000			
700	277	90	62	18	—	160	229.6	c
5000	389	113	436	"	—	567	113.4	c
50000	1309	380	4357	"	—	4755	95.1	c
100000	2137	620	8714	"	—	9352	93.5	c
700	319	93	35	22	—	150	214.3	c
5000	476	138	258	"	—	418	83.6	c
50000	1676	486	2582	"	—	3090	61.8	c
100000	2793	810	5163	"	—	5995	59.9	c
700	343	99	25	27	—	151	215.7	c
5000	537	156	174	"	—	357	71.4	c
50000	1795	579	1743	"	—	2349	47.0	c
100000	3591	1041	3486	"	—	4554	45.5	c
700	343	99	14	32	—	145	207.1	c
5000	592	172	97	"	—	301	60.2	c
50000	2860	829	968	"	—	1829	36.6	c
100000	5187	1504	1935	"	—	3471	34.7	c
700	360	101	62	36	38	161	230.0	c
5000	613	178	438	"	272	380	76.0	c
50000	2625	761	4378	"	2717	2458	49.2	c
100000	4465	1295	8756	"	5435	4652	46.5	c
700	335	97	62	"	38	157	224.3	c
5000	580	168	438	"	272	370	74.0	c
50000	2417	701	4378	"	2717	2398	48.0	c
100000	4132	1198	8756	"	5435	4655	45.6	c

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OFF GAS HEAT CONTENT _____ BTU/SCF
 RESIDENCE TIME 75 SEC.
 COMB. TEMP 1600 °F

OFF GAS FLOW	CAPITAL COST	FIXED COSTS	UTILITIES	MANPOWER	RECOVERY CREDIT	NET ANNUALIZED COST - OR - SAVINGS	NET COST OR SAVING PER CUBIC FOOT
SCFM	\$1000	\$1000	\$1000	\$1000	\$1000	\$1000	\$/SCFM
700	284	82	62	18	—	162	231.4
5000	429	125	436	"	—	579	115.7
50000	1692	491	4357	"	—	4866	97.3
100000	2817	817	8714	"	—	9549	95.5
700	329	75	35	22	—	152	217.1
5000	523	152	258	"	—	432	86.4
50000	1995	579	2582	"	—	3183	63.7
100000	3392	984	5163	"	—	6169	61.7
700	346	100	25	27	—	152	217.1
5000	567	164	174	"	—	365	73.0
50000	2391	679	1743	"	—	2449	49.0
100000	4150	1204	3486	"	—	4717	47.2
700	352	102	14	32	—	148	211.4
5000	609	177	97	"	—	306	61.2
50000	3192	926	968	"	—	1926	38.5
100000	5613	1628	1935	"	—	3595	36.0
700	349	101	62	36	38	161	230.0
5000	655	190	438	"	272	392	78.4
50000	3009	873	4378	"	2717	2570	51.4
100000	5145	1492	8756	"	5435	4849	48.5
700	331	96	62	"	38	156	222.9
5000	621	180	438	"	272	382	76.4
50000	2800	812	4378	"	2717	2509	50.2
100000	4812	1395	8756	"	5435	4752	47.5

OFF GAS HEAT CONTENT 10 BTU/SCF
 RESIDENCE TIME 0.5 SEC.
 COMB. TEMP 1600 °F

OFF GAS FLOW	SCFM	CAPITAL COST \$1000	OPERATING COSTS-OR- CREDIT				NET ANNUALIZED COST-OR- SAVINGS \$1000	NET COST OR SAVING PER SCFM \$/SCFM	C
			FIXED COSTS \$1000	UTILITIES \$1000	MANPOWER \$1000	RECOVERY CREDIT \$1000			
11-20 HEAT RECOVERY	700	277	80	55	18	—	153	218.6	C
	5000	389	113	387	"	—	518	103.5	C
	50000	1309	380	3868	"	—	4256	85.1	C
	100000	2137	620	7736	"	—	8374	83.7	C
11-20 HEAT RECOVERY	700	319	93	29	22	—	144	205.7	C
	5000	476	138	207	"	—	367	73.4	C
	50000	1,676	486	2,064	"	—	2,572	51.4	C
	100000	2,793	810	4,127	"	—	4,959	49.6	C
11-50 HEAT RECOVERY	700	343	99	17	27	—	143	204.3	C
	5000	537	156	118	"	—	301	60.2	C
	50000	1,995	579	1,182	"	—	1,788	36.8	C
	100000	3,591	1,041	2,365	"	—	3,433	34.3	C
11-70 HEAT RECOVERY	700	343	99	5	32	—	136	194.3	C
	5000	592	172	37	"	—	241	48.2	C
	50000	2,860	829	372	"	—	1,233	24.7	C
	100000	5,187	1,504	743	"	—	2,279	22.8	C
11-400 WASTE HT. BOILER RECOVERY SYSTEM	700	350	101	55	36	38	154	220.0	C
	5000	613	178	389	"	270	331	66.3	C
	50000	2,625	761	3889	"	2717	1,969	39.4	C
	100000	4,465	1,245	7778	"	5435	3,674	36.7	C
11-250 WASTE HT. BOILER RECOVERY SYSTEM	700	335	97	65	"	38	160	214.3	C
	5000	580	168	389	"	270	321	64.3	C
	50000	2,417	701	3889	"	2717	1,909	38.3	C
	100000	4,132	1,198	7779	"	5435	3,577	35.8	C

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OFF GAS HEAT CONTENT 10 BTU/SCF
RESIDENCE TIME .75 SEC.
COLAB. TEMP 1600 °F

Case	OFF GAS FLOW SCFM	CAPITAL COST \$1000	OPERATING COSTS-OR- CREDIT			NET ANNUALIZED COST - OR - SAVINGS \$1000	NET COST OR SAVING PER YEAR \$/SCFM
			FIXED COSTS \$1000	UTILITIES \$1000	MANPOWER \$1000	RECOVERY CREDIT \$1000	
I 10% HEAT RECOVERY	700	384	82	55	18	—	231.4
	5000	424	125	387	"	—	106.9
	50000	1692	491	3868	"	—	87.5
	100000	2817	817	7736	"	—	85.7
II-33 30% HEAT RECOVERY	700	329	95	29	22	—	208.6
	5000	523	152	207	"	—	76.2
	50000	1995	579	2064	"	—	53.3
	100000	3392	984	4127	"	—	51.3
II-33 50% HEAT RECOVERY	700	346	100	17	27	—	205.7
	5000	567	164	118	"	—	61.8
	50000	2341	679	1182	"	—	37.8
	100000	4150	1204	2365	"	—	36.0
II-73 10% HEAT RECOVERY	700	352	102	5	32	—	198.6
	5000	609	177	37	"	—	49.2
	50000	3192	926	372	"	—	26.6
	100000	5613	1628	743	"	—	24.0
III-100 10% HEAT RECOVERY	700	349	101	55	36	38	220.0
	5000	655	190	389	"	272	68.6
	50000	3009	873	3889	"	2717	41.6
	100000	5145	1492	7778	"	5435	38.7
III-200 10% HEAT RECOVERY	700	331	96	65	"	38	212.9
	5000	621	180	389	"	272	66.6
	50000	2800	812	3889	"	2717	40.4
	100000	4812	1395	7778	"	5435	37.7

OFF GAS HEAT CONTENT 13 BTU/SCF
 RESIDENCE TIME 0.5 SEC.
 COMB. TEMP 1600 °F

CASE	OFF GAS FLOW SCFM	CAPITAL COST \$1000	OPERATING COSTS-OR-CREDIT			NET ANNUALIZED COST - OR - SAVINGS \$1000	NET COST OR SAVING PER SCFM \$/SCFM	C/S
			FIXED COSTS \$1000	UTILITIES \$1000	MANPOWER \$1000	RECOVERY CREDIT \$1000		
I NO HEAT RECOVERY	700	277	80	52	18	—	214.3	C
	5000	388	113	371	"	—	100.3	C
	50000	1309	380	2705	"	—	82.1	C
	100000	2137	620	7411	"	—	80.5	C
II-30 30% HEAT RECOVERY <i>not provided</i>	700	319	93	26	22	—	201.4	C
	5000	476	138	187	"	—	69.4	C
	50000	1676	486	1873	"	—	47.6	C
	100000	2793	810	3745	"	—	45.8	C
II-50 50% HEAT RECOVERY <i>not provided</i>	700	343	99	15	27	—	201.4	C
	5000	537	156	100	"	—	56.6	C
	50000	1995	579	999	"	—	32.1	C
	100000	3591	1041	1998	"	—	30.7	C
II-70 70% HEAT RECOVERY <i>not provided</i>	700	343	99	3	32	—	191.4	C
	5000	592	172	18	"	—	44.4	C
	50000	2860	829	178	"	—	20.8	C
	100000	5187	1504	356	"	—	18.9	C
III-100 WASTE HT. BOILER ADDS 100 PSI STIA.	700	360	101	52	36	38	215.7	C
	5000	613	178	373	"	272	63.0	C
	50000	2625	761	3726	"	2717	36.1	C
	100000	4465	1295	7452	"	5435	33.5	C
III-250 WASTE HT. BOILER 250 PSI STIA.	700	335	97	52	"	38	210.0	C
	5000	580	168	373	"	272	61.0	C
	50000	2417	701	3726	"	2717	34.9	C
	100000	4132	1198	7452	"	5435	32.5	C

OFF GAS HEAT CONTENT 13 BTU/SCF
 RESIDENCE TIME 75 SEC.
 COMB. TEMP 1600 °F

CASE	OFF GAS FLOW SCFM	CAPITAL COST \$1000	OPERATING COSTS-OR- CREDIT			NET ANNUALIZED COST - OR - SAVINGS \$/1000	NET COST OR SAVING PER SCFM \$/SCFM	%
			FIXED COSTS \$1000	UTILITIES \$1000	MANPOWER \$1000	RECOVERY CREDIT \$/1000		
I NO HEAT RECOVERY	700	284	82	52	18	—	214.3	C
	5000	429	125	371	"	—	102.7	C
	50000	1692	491	3705	"	—	84.3	C
	100000	2817	817	7411	"	—	82.5	C
II-30 30% HEAT RECOVERY <i>assumed</i>	700	329	95	26	22	—	204.3	C
	5000	523	152	187	"	—	72.2	C
	50000	1995	579	1873	"	—	49.5	C
	100000	3392	984	3745	"	—	47.5	C
II-50 50% HEAT RECOVERY <i>assumed</i>	700	346	100	15	27	—	202.9	C
	5000	567	164	100	"	—	58.2	C
	50000	2341	679	999	"	—	34.1	C
	100000	4150	1204	1998	"	—	32.3	C
II-70 70% HEAT RECOVERY <i>assumed</i>	700	352	102	3	32	—	195.7	C
	5000	609	177	18	"	—	45.4	C
	50000	3192	926	178	"	—	22.7	C
	100000	5613	1628	356	"	—	20.2	C
III-400 WASTE HT. BOILER 400PSI STM.	700	349	101	52	36	38	215.7	C
	5000	655	190	373	"	272	65.4	C
	50000	3009	873	3726	"	2717	38.4	C
	100000	5145	1492	7452	"	5436	35.5	C
III-250 WASTE HT. BOILER 250PSI STM.	700	331	96	52	"	38	208.6	C
	5000	621	180	373	"	272	63.4	C
	50000	2800	812	3726	"	2717	37.1	C
	100000	4812	1395	7452	"	5435	34.5	C

OFF GAS HEAT CONTENT 20 BTU/SCF
 RESIDENCE TIME 0.5 SEC.
 COMB. TEMP 1600 °F

CASE	OFF GAS FLOW SCFM	CAPITAL COST \$1000	OPERATING COSTS-OR-CREDIT				NET ANNUALIZED COST-OR- SAVINGS \$1000	NET COST OR SAVING PER SCFM \$/SCFM	15/5
			FIXED COSTS \$1000	UTILITIES \$1000	MANPOWER \$1000	RECOVERY CREDIT \$1000			
I 100% HEAT RECOVERY	700	277	80	46	18	—	144	205.7	C
	5000	388	113	332	"	—	463	92.6	C
	50000	1309	380	3316	"	—	3714	74.3	C
	100000	2137	620	6633	"	—	7271	72.7	C
II-30 30% HEAT RECOVERY	700	319	93	17	22	—	132	188.6	C
	5000	476	138	129	"	—	289	57.8	C
	50000	1676	486	1291	"	—	1799	26.0	C
	100000	2793	810	2579	"	—	3411	24.1	C
II-50 50% HEAT RECOVERY	700	343	99	7	27	—	133	190.0	C
	5000	537	156	61	"	—	244	48.8	C
	50000	1995	579	609	"	—	1215	29.3	C
	100000	3591	1091	1218	"	—	2286	22.9	C
II-70 70% HEAT RECOVERY	700								
	5000								
	50000								
	100000								
III-100 WASTE HT. BOILER 400 PSI. STIA.	700	350	101	46	36	38	145	207.1	C
	5000	613	178	334	"	272	276	65.2	C
	50000	2625	761	3337	"	2717	1417	28.3	C
	100000	4465	1295	6675	"	5435	2571	25.7	C
III-250 WASTE HT. BOILER 250 PSI. STIA.	700	335	97	46	"	38	141	201.4	C
	5000	580	168	334	"	272	266	53.2	C
	50000	2417	701	3337	"	2717	1357	27.1	C
	100000	4132	1198	6675	"	5435	2474	24.7	C

OFF GAS HEAT CONTENT 20 BTU/SCF
 RESIDENCE TIME 7.5 SEC.
 COMB. TEMP 1600 °F

CASE	OFF GAS FLOW SCFM	CAPITAL COST \$1000	OPERATING COSTS-OR- CREDIT				NET ANNUALIZED COST - OR - SAVINGS \$1000	NET COST OR SAVING PER SCFM \$/SCFM	%
			FIXED COSTS \$1000	UTILITIES \$1000	MANPOWER \$1000	RECOVERY CREDIT \$1000			
I NO HEAT RECOVERY	700	284	82	46	18	—	146	208.6	C
	5000	429	125	332	"	—	475	94.9	C
	50000	1692	491	3316	"	—	3825	76.5	C
	100,000	2817	817	6633	"	—	7468	74.7	C
II-30 30% HEAT RECOVERY	700	329	95	17	23	—	134	191.4	C
	5000	523	152	129	"	—	303	60.6	C
	50000	1,995	571	1,291	"	—	1,892	37.8	C
	100,000	3,392	984	2,579	"	—	3,585	35.9	C
II-50 50% HEAT RECOVERY	700	346	100	7	27	—	134	191.4	C
	5000	567	164	61	"	—	252	50.4	C
	50000	2,341	679	609	"	—	1,315	26.3	C
	100,000	4,150	1,204	1,418	"	—	2,449	24.5	C
II-70 70% HEAT RECOVERY	700	—	—	—	—	—	—	—	—
	5000	—	—	—	—	—	—	—	—
	50000	—	—	—	—	—	—	—	—
	100,000	—	—	—	—	—	—	—	—
III-400 WASTE HT. BOILER 400PSI. STM.	700	349	101	46	36	38	145	202.1	C
	5000	655	190	334	"	272	288	57.6	C
	50000	3009	873	3337	"	2717	1529	30.6	C
	100,000	5145	1492	6675	"	5435	2768	27.7	C
III-250 WASTE HT. BOILER 250PSI. STM.	700	331	96	46	"	38	140	200.0	C
	5000	621	180	334	"	272	273	54.6	C
	50000	2800	812	3337	"	2717	1468	29.4	C
	100,000	4812	1395	6675	"	5435	2671	26.7	C

OFF GAS HEAT CONTENT 50 BTU/SCF
 RESIDENCE TIME 0.5 SEC.
 COMB. TEMP 1600 °F

CASE	OFF GAS FLOW SCFM	CAPITAL COST \$1000	OPERATING COSTS-OR- CREDIT			NET ANNUALIZED COST - OR - SAVINGS \$1000	NET COST OR SAVING PER SCFM \$/SCFM	%
			FIXED COSTS \$1000	UTILITIES \$1000	MANPOWER \$1000			
I NO HEAT RECOVERY	700	277	80	24	18	122	174.3	C
	5000	388	113	167	"	298	59.6	C
	50000	1309	380	1667	"	2065	41.3	C
	100000	2137	620	3332	"	3970	39.7	C
II-30 30% HEAT RECOVERY	700							
	5000							
	50000							
	100000							
II-50 50% HEAT RECOVERY	700							
	5000							
	50000							
	100000							
II-70 70% HEAT RECOVERY	700							
	5000							
	50000							
	100000							
III-400 WASTE HT. BOILER 400PSI. STM.	700	350	101	24	36	123	175.7	C
	5000	613	178	169	"	111	22.2	C
	50000	2625	761	1687	"	233	4.7	S
	100000	4465	1295	3374	"	730	7.3	S
III-250 WASTE HT. BOILER 250PSI. STM.	700	335	97	24	"	114	170.0	C
	5000	580	168	169	"	101	20.2	C
	50000	2417	701	1687	"	293	5.9	S
	100000	4132	1198	3374	"	827	8.4	S

OFF GAS HEAT CONTENT 50 BTU/SCF
 RESIDENCE TIME 75 SEC.
 COMB. TEMP 1600 °F

GAS	OFF GAS FLOW SCFM	CAPITAL COST \$1000	OPERATING COSTS-OR-CREDIT				NET ANNUALIZED COST-OR- SAVINGS \$1000	NET COST OR SAVING PER SCFM \$/SCFM	C/S
			FIXED COSTS \$1000	UTILITIES \$1000	MANPOWER \$1000	RECOVERY CREDIT \$1000			
I 10% HEAT RECOVERY	700	284	82	24	18	—	124	177.1	C
	5000	429	125	167	"	—	310	62.0	C
	50000	1692	491	1667	"	—	2176	43.5	C
	100000	2817	817	3332	"	—	4167	41.7	C
II-30 30% HEAT RECOVERY	700								
	5000								
	50000								
	100000								
II-50 50% HEAT RECOVERY	700								
	5000								
	50000								
	100000								
II-70 70% HEAT RECOVERY	700								
	5000								
	50000								
	100000								
III-100 WASTE HT. BOILER STHA.	700	349	101	24	36	38	123	175.7	C
	5000	655	190	169	"	272	123	24.6	C
	50000	3009	873	1687	"	2717	121	2.4	S
	100000	5145	1492	3374	"	5435	533	5.3	S
III-250 WASTE HT. BOILER STHA.	700	331	96	24	"	38	118	168.6	C
	5000	621	180	169	"	272	113	22.6	C
	50000	2800	812	1687	"	2717	182	3.6	S
	100000	4812	1395	3374	"	5435	630	6.3	S

OFF GAS HEAT CONTENT 100 BTU/SCF

RESIDENCE TIME 5 SEC.

COMB. TEMP 1815 °F

CASE	OFF GAS FLOW SCFM	CAPITAL COST \$1000	OPERATING COSTS-OR-CREDIT			NET ANNUALIZED COST-OR- SAVINGS \$1000	NET COST OP. SAVING PER SCFM \$/SCFM	C
			FIXED COSTS \$1000	UTILITIES \$1000	MANPOWER \$1000	RECOVERY CREDIT \$1000		
I 100% HEAT RECOVERY	500	267	83	7	18	-	204.3	C
	5000	440	129	73	19	-	43.7	
	50000	1651	471	731	19	-	24.6	
	100000	2949	826	1462	19	-	23.1	
II-30 30% HEAT RECOVERY	500							
	5000							
	50000							
	100000							
II-50 50% HEAT RECOVERY	500							
	5000							
	50000							
	100000							
II-70 70% HEAT RECOVERY	500							
	5000							
	50000							
	100000							
III-100 WASTE HT. BOILER 400PSI ST.M.	500	381	111	8	36	43	223.2	S
	5000	801	232	76	36	429	169	S
	50000	3607	1046	766	36	4290	49.0	S
	100000							
III-250 WASTE HT. BOILER 250PSI ST.M.	500	365	106	8	36	43	213.9	C
	5000	744	216	76	36	429	20.2	S
	50000	3291	954	766	36	4290	50.9	S
	100000							

OFF GAS HEAT CONTENT 100 BTU/SCF
 RESIDENCE TIME .75 SEC.
 COMB. TEMP 1975 °F

CASE	OFF GAS FLOW SCFM	CAPITAL COST \$1000	OPERATING COSTS-OR- CREDIT			NET ANNUALIZED COST - OR - SAVINGS \$1000	NET COST OR SAVING PER SCFM
			FIXED COSTS \$1000	UTILITIES \$1000	MANPOWER \$1000		
I 10% HEAT RECOVERY	500	292	85	7	18	110	219.5
	5000	509	149	73	18	239	477
	50000	2151	624	731	18	1373	275
	100000	3876	1124	1462	18	2604	260
II-30 30% HEAT RECOVERY	500						
	5000						
	50000						
	100000						
II-50 50% HEAT RECOVERY	500						
	5000						
	50000						
	100000						
II-70 70% HEAT RECOVERY	500						
	5000						
	50000						
	100000						
III-100 WASTE HT. BOILER 600PSI STM.	500	387	112	8	36	113	2263
	5000	872	253	76	36	64	128
	50000	4107	1191	756	36	2307	462
	100000	7576	1512	1512	36		
III-250 WASTE HT. BOILER 250PSI STM.	500	371	107	9	36	108	2120
	5000	813	236	76	36	81	162
	50000	3791	1099	756	36	2399	480
	100000	6979	1512	1512	36		

OFF GAS HEAT CONTENT 200 BTU/SCF
 RESIDENCE TIME 15 SEC.
 COMB. TEMP 2200 °F

CASE	OFF GAS FLOW SCFM	CAPITAL COST \$1000	OPERATING COSTS-OR-CREDIT				NET ANNUALIZED COST-OR- SAVINGS \$1000	NET COST OR SAVING PER SCFM \$/SCFM	C/S
			FIXED COSTS \$1000	UTILITIES \$1000	MANPOWER \$1000	RECOVERY CREDIT \$1000			
I-10 HEAT RECOVERY	500	249	67	11	18	-	114	23.12	C
	5000	557	162	113	18	-	293	5.85	C
	50000	2629	762	1129	18	-	1909	38.2	C
	100000	-	-	-	-	-	-	-	-
II-30 30% HEAT RECOVERY	500	-	-	-	-	-	-	-	-
	5000	-	-	-	-	-	-	-	-
	50000	-	-	-	-	-	-	-	-
	100000	-	-	-	-	-	-	-	-
II-50 50% HEAT RECOVERY	500	-	-	-	-	-	-	-	-
	5000	-	-	-	-	-	-	-	-
	50000	-	-	-	-	-	-	-	-
	100000	-	-	-	-	-	-	-	-
II-70 70% HEAT RECOVERY	500	-	-	-	-	-	-	-	-
	5000	-	-	-	-	-	-	-	-
	50000	-	-	-	-	-	-	-	-
	100000	-	-	-	-	-	-	-	-
III-400 WASTE HT. BOILER 400PSI. STM.	500	407	118	12	36	86	80	15.89	C
	5000	1180	342	117	36	459	364	7.28	S
	50000	6151	1784	1171	36	8587	5596	111.9	S
	100000	-	-	-	-	-	-	-	-
III-250 WASTE HT. BOILER 250PSI. STM.	500	389	113	12	36	86	75	14.78	C
	5000	1072	311	117	36	859	34	7.90	S
	50000	5580	1618	1171	36	8597	5762	115.2	S
	100000	-	-	-	-	-	-	-	-

OFF GAS HEAT CONTENT 200 BTU/SCF
 RESIDENCE TIME .75 SEC.
 COMB. TEMP 2200 °F

CASE	OFF GAS FLOW SCFM	CAPITAL COST \$1000	OPERATING COSTS-OR-CREDIT				NET ANNUALIZED COST-OR-SAVINGS \$1000	NET COST OR SAVING PER SCFM \$/SCFM
			FIXED COSTS \$1000	UTILITIES \$1000	MANPOWER \$1000	RECOVERY CREDIT \$1000		
I NO HEAT RECOVERY	500	313	91	11	19	-	120	23.91
	5000	668	194	113	19	-	325	6.49
	50000	3431	995	1129	19	-	2192	4.28
	100,000	-	-	-	-	-	-	-
II-30 30% HEAT RECOVERY	500	-	-	-	-	-	-	-
	5000	-	-	-	-	-	-	-
	50000	-	-	-	-	-	-	-
	100,000	-	-	-	-	-	-	-
II-50 50% HEAT RECOVERY	500	-	-	-	-	-	-	-
	5000	-	-	-	-	-	-	-
	50000	-	-	-	-	-	-	-
	100,000	-	-	-	-	-	-	-
II-70 70% HEAT RECOVERY	500	-	-	-	-	-	-	-
	5000	-	-	-	-	-	-	-
	50000	-	-	-	-	-	-	-
	100,000	-	-	-	-	-	-	-
III-100 WASTE HT. BOILER 600 PSI STIA.	500	431	122	12	36	86	84	16.89
	5000	1242	375	117	36	859	331	6.63
	50000	6953	2016	1171	36	8587	5364	10.73
	100,000	-	-	-	-	-	-	-
III-250 WASTE HT. BOILER 250 PSI STIA.	500	404	117	12	36	86	79	15.83
	5000	1184	343	117	36	859	363	7.25
	50000	6384	1851	1171	36	8587	5529	11.05
	100,000	-	-	-	-	-	-	-

REPORT 2

CONTROL DEVICE EVALUATION
THERMAL OXIDATION SUPPLEMENT
(VOC CONTAINING HALOGENS OR SULFUR)

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November 1980

D80K

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I. INTRODUCTION

This report is a supplement to the control device evaluation report for thermal oxidation* authored by J. W. Blackburn, July 1980, and describes the control technology and costs for two additional cases. In the first case conventional thermal oxidation is used (combustion temperature of 1200 to 1600°F) but, due to the presence of sulfur-containing VOC in the waste gas, flue-gas scrubbing is required for removal of SO₂. The second case is for control of halogen-containing VOC in the waste, which requires high-temperature thermal oxidation (combustion temperatures of 1800 to 3000°F) and flue-gas scrubbing. The flue-gas scrubber design is assumed to be similar in each application and to be an additional cost. The cost of the conventional thermal oxidizers for the first case is assumed to be the same as that given in the control-device evaluation report for thermal oxidation whereas a new set of cost data was developed for the thermal oxidizers in the second case, which also included the cost for a waste-heat steam-generation boiler.

Conventional thermal oxidation units for VOC control of non-sulfur-containing and non-halogen-containing compounds normally have combustion chambers with temperatures of 1400 or 1600°F for low-heat-content gases and temperatures in the 2200°F range for high-heat-content (100 Btu/scf) gases. The heat content is a measure of how much heat is generated by the gas during combustion and is determined by the VOC concentration and its heat of combustion. The heat content and the waste-gas flow determine the combustion chamber design and auxiliary fuel usage. The flue-gas exhaust is usually vented to the atmosphere through a stack without further treatment, although heat recovery units such as recuperative heaters or waste-heat steam-generation boilers are installed when there is an economic incentive. The technical analysis of conventional thermal oxidizers is included in the thermal oxidation report.^{1*}

Waste gases containing sulfur or halogen compounds require flue-gas scrubbing after thermal oxidation to remove the noxious gases that are formed during combustion. The flue-gas exhaust from the combustion chamber is first sent into a water quench chamber to be cooled to its adiabatic saturation

*See References, Sect. VII.

temperature and is then routed through the scrubber to remove the noxious gases. After the flue gas goes through the scrubber, it is vented to the atmosphere through a stack. Waste-heat recovery units can be installed before the water quench chamber when heat recovery is desired. Figure I-1 is a flow diagram of a typical thermal oxidation control device with a scrubber.

Thermal oxidation control of sulfur-containing VOC can be achieved with the thermal oxidation design procedure discussed in the thermal oxidation report for non-sulfur-containing and non-halogen-containing VOC control, except that a quench chamber and scrubber must be provided to remove the noxious gases produced during combustion. In this report the design and cost of the quench chamber and scrubber unit are given, together with operating costs associated with the control of sulfur-containing VOC. The thermal oxidation system costs were obtained from the control device evaluation report for thermal oxidation.

Thermal oxidation control of halogen-containing VOC requires high-temperature oxidation to convert the combustion product to a form that can most easily be removed by scrubbing. For instance, chloride-containing waste gases are burned at high temperature to convert the chlorine to HCl instead of to Cl_2 , since HCl is more easily scrubbed. In this report the design and cost of thermal oxidation systems for combustion temperatures of 1800 to 3000°F are presented, along with the design and cost of the quench chamber and scrubber required in this temperature range. The waste-heat steam-generation boiler costs for high-temperature thermal oxidation are also included. The cost of a thermal oxidizer and waste-heat steam-generation boiler for a combustion temperature of 1800 and 2200°F is obtained from in the control device evaluation report for thermal oxidation.

The heat content range of the waste gas used in this report is 1 to 100 Btu/scf. For waste gases with heat contents of less than 100 Btu/scf, supplementary fuel must be added to maintain a combustion temperature above 2200°F. Gases with heat contents above 100 Btu/scf were not considered since most sulfur-containing and halogen-containing compounds will not have a higher heat content unless a very high concentration is achieved. Heat contents of 13 and 20 Btu/scf in air normally correspond for most compounds to 25 and 40% of the lower explosive limits (LEL). Waste gases with heat contents of 20 to 50 Btu/scf (approximately 40 to 100% of the LEL) must be diluted with air or be enriched with

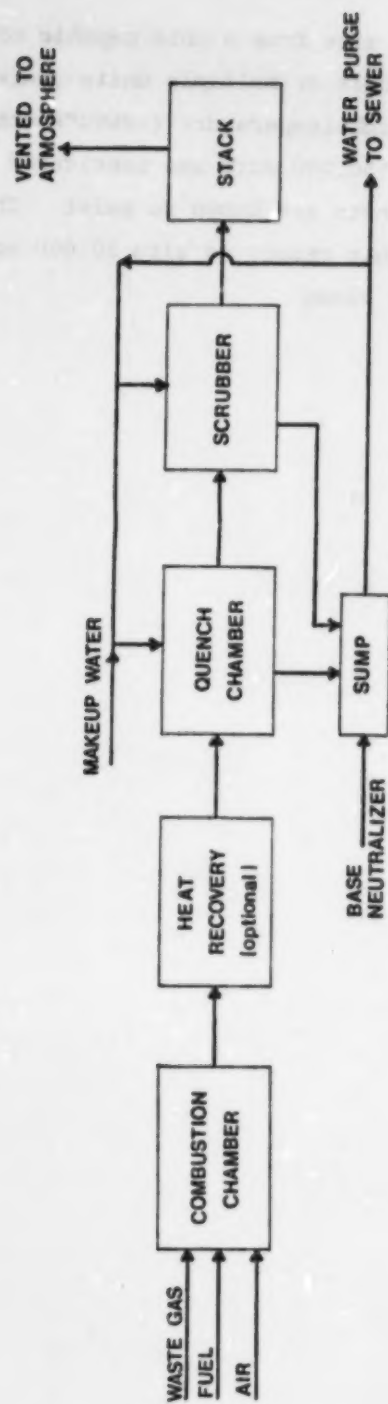


Fig. I-1. Process Flowsheet for Thermal Oxidation with a Flue-Gas Scrubber

auxiliary fuel because they exceed the flammable safety limits imposed by insurance companies.

Conventional thermal oxidizers range in size from a unit capable of controlling several hundred scfm of waste gas to single or multiple units controlling waste gases in excess of 100,000 scfm. For high-temperature ($>2600^{\circ}\text{F}$) thermal oxidation units an upper size limit of $\sim 50,000$ scfm was considered to be reasonable for this report since no 100,000-scfm units are known to exist. The upper size limit for the scrubber design used in this report is also 50,000 scfm. Multiple scrubbers would be required for greater flows.

II. THERMAL OXIDIZER AND SCRUBBER DESIGN CONSIDERATIONS

The first report on thermal oxidation discussed the effects that the combustion chamber residence time and temperature have on the design for thermal oxidizers in the range of 1200 to 2200°F. In this temperature range the residence time and temperature affect the destruction efficiency of the VOC. The temperature also affects the auxiliary heat required for flame stability, the method used for heat recovery, and the capital and operating costs of the system. For control of sulfur-containing VOC the same considerations and design criteria for thermal oxidizers hold true except that a scrubber is used after the combustion chamber to control SO₂ emissions.

The design considerations and assumptions used in this report are that halogen-containing VOC are controlled by high-temperature thermal oxidation, that a scrubbing system is used for high-temperature and conventional systems with noxious flue gases, and that consideration must be given to heat recovery for conventional and high-temperature systems.

A. HIGH-TEMPERATURE THERMAL OXIDATION

For the control of halogenated VOC a higher temperature is required to convert the noxious combustion products to a more easily removed form. The halogen-containing VOC most commonly encountered is the chlorinated VOC, which is the basis of the discussion and economics in this report. Bromine-containing VOC are more difficult to control due to the more severe conditions required to convert Br₂ to HBr. Fluorine-containing VOC are more difficult to control due to the high corrosiveness of HF; however, high-temperatures are not required.

A combustion chamber temperature of above 1800°F for a well-designed oxidizer with a proper residence time will have essentially 99.9% VOC destruction efficiency.^{2,3} The residence time in the combustion chamber is a design variable specified by the system designer. The combustion chamber is a chemical reactor, and the residence time is the time available for the reaction (oxidation) to occur. Residence times as low as 0.3 sec to several seconds have been utilized in thermal oxidizer designs. The fraction of the total chamber volume that is at the combustion temperature depends on the chamber configuration and the design of the flame burner.

Although a design engineer should be quite concerned with the actual burner design and residence time at the combustion temperature, in this control-device evaluation study it is assumed that the entire combustion chamber volume is at the combustion temperature. This assumption is justified since later it is shown that differences as high as 50% in residence time do not significantly affect the annual cost of thermal oxidation control; however, the capital cost is affected.

The need of going to a higher combustion temperature for chlorinated compounds is best shown by the relationship of the reaction of Cl_2 and water.⁴ Figure II-1 shows the effect of combustion temperature on the equilibrium constant for the following reaction:



The equilibrium constant (K_p) indicates the relative concentration of Cl_2 and HCl in the exhaust gas from the combustion chamber by the equation

$$K_p = \frac{(\text{HCl})^2 (\text{O}_2)^{\frac{1}{2}}}{(\text{Cl}_2) (\text{H}_2\text{O})}$$

The higher the value of K_p , the more HCl will be formed, which is easier to remove from the exhaust gas than Cl_2 . At low temperatures ($<2000^\circ\text{F}$) the equilibrium constant decreases rapidly, indicating that the Cl_2 concentration in the exhaust will increase appreciably. To achieve an acceptable low Cl_2 concentration in the flue gas, temperatures of above 1800°F are necessary. Another way to increase the conversion of Cl_2 to HCl is to inject steam into the combustion chamber to force the equilibrium toward HCl . The actual flue-gas concentration of Cl_2 will not directly follow the value indicated by Fig. II-1 since the kinetics of the reaction may prevent equilibrium from being achieved in the residence time available.⁵ The reverse reaction may occur as the combustion products are cooled. The reverse reaction can be minimized by very rapid cooling but this may prevent effective heat recovery.⁵ Selection of the appropriate combustion temperature requires optimization of the operating costs with the capital cost involved for the conversion. An optimum design will differ for different chlorinated feeds. In this report it is assumed that the

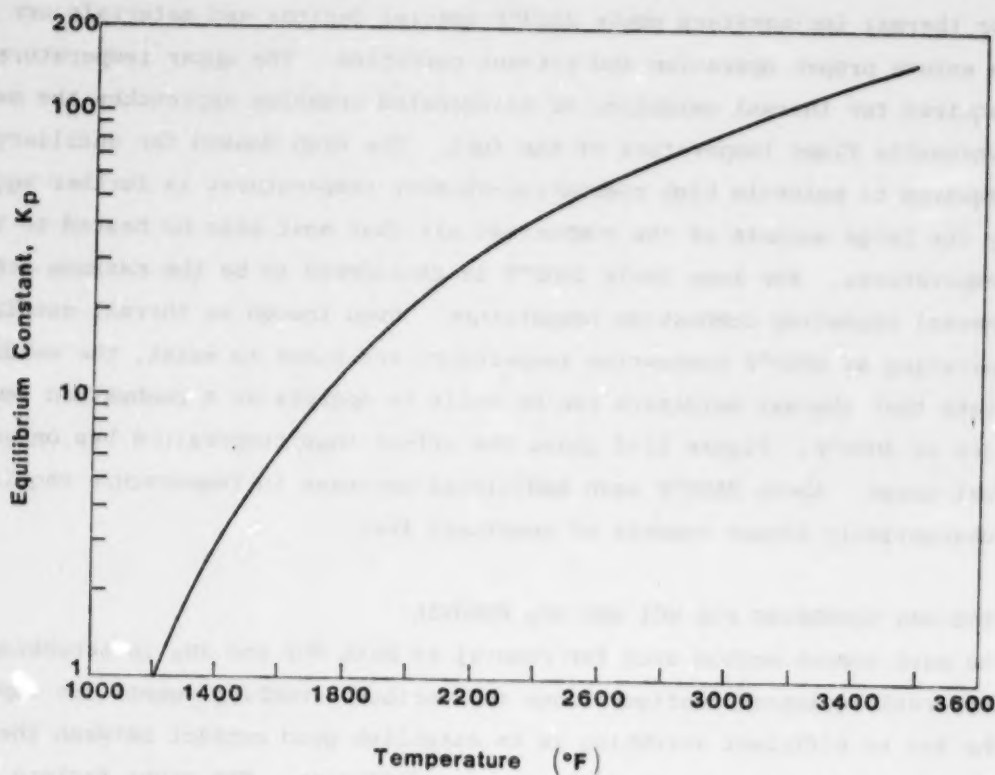


Fig. II-1. Cl_2/HCl Equilibrium Constant as a Function of Temperature
for the Reaction $\text{Cl}_2 + \text{H}_2\text{O} \rightleftharpoons 2\text{HCl} + 1/2\text{O}_2$

VOC in the waste gas is 100% halogenated. When the halogen-containing VOC is actually a small percentage of the VOC in the waste gas, the optimum combustion temperature will differ.

For thermal incinerators above 2200°F special designs and materials are required to ensure proper operation and prevent corrosion. The upper temperature range required for thermal oxidation of halogenated organics approaches the maximum achievable flame temperature of the fuel. The high demand for auxiliary fuel required to maintain high combustion-chamber temperatures is further aggravated by the large amounts of the combustion air that must also be heated to these temperatures. For some fuels 3000°F is considered to be the maximum achievable thermal oxidation combustion temperature. Even though no thermal oxidizers operating at 3000°F combustion temperature are known to exist, the vendors state that thermal oxidizers can be built to operate at a combustion temperature of 3000°F. Figure II-2 shows the effect that temperature has on auxiliary fuel usage. Above 2600°F each additional increase in temperature requires substantially larger amounts of auxiliary fuel.

B. FLUE-GAS SCRUBBING FOR HCl AND SO₂ REMOVAL

The most common method used for removal of both HCl and SO₂ is scrubbing. Many different equipment configurations and various scrubbing agents are employed. The key to efficient scrubbing is to establish good contact between the gas and the liquid to effect complete interphase diffusion. The other factors affecting scrubbing efficiency are the temperature, pH, and alkalinity of the scrubbing agent.

The flue-gas scrubbing operation involves three basic steps: quenching of the flue gas, contacting the flue gas with the scrubbing agent, and separating the scrubbed material from the scrubbing agent. Quenching of the flue gas will reduce the temperature of the flue gas and saturate it with water. Some of the SO₂ or HCl will be removed from the flue gas during the quenching operation. The configuration and operation of the flue-gas scrubbing-agent contacting equipment will determine the efficiency of the SO₂ and HCl removal from the flue gas. The method used to separate the SO₂ or HCl from the scrubbing agent will determine the final disposal of the sulfur and chlorine compounds. The methods used range from purging the scrubbing agent and discharging it to the sewer to recovery of the scrubbing agent for reuse.

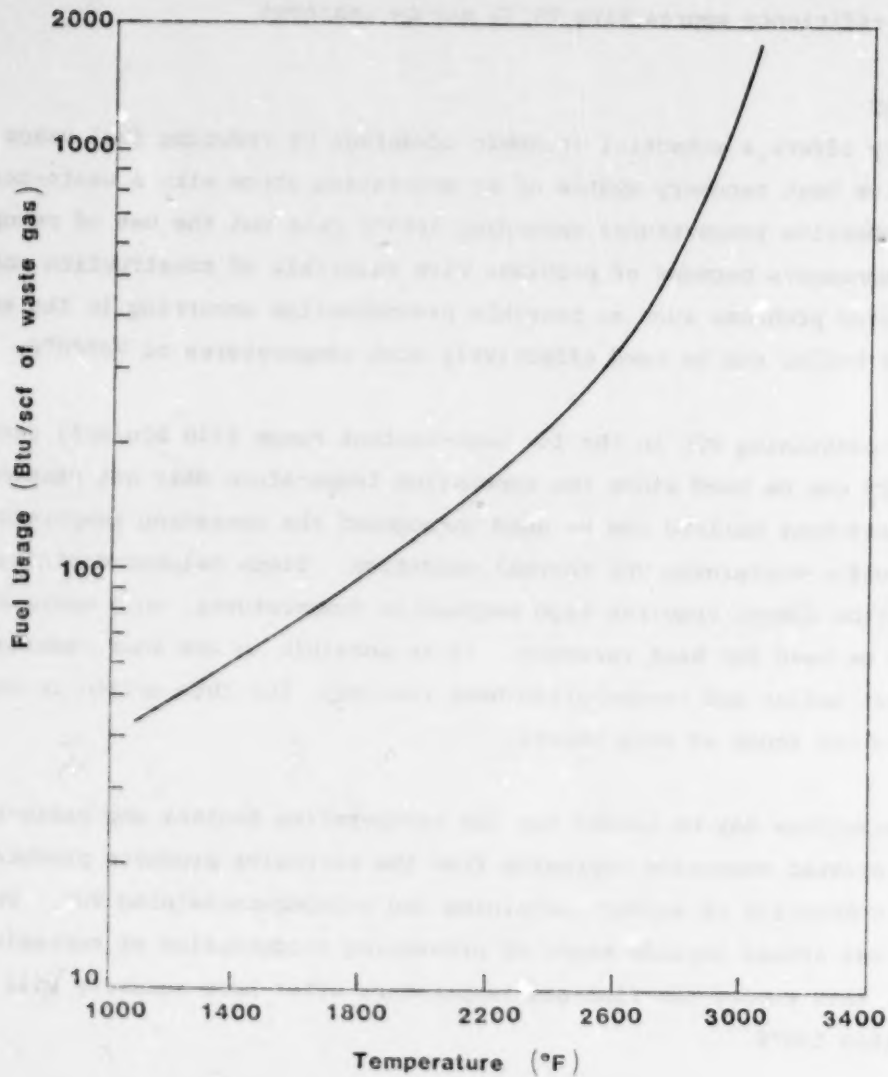


Fig. II-2. Temperature vs Fuel Usage for a Waste Gas at a Heat Content of 1 Btu/scf with 3 mole % O_2 in the Flue Gas

The efficiency achieved for SO_2 removal by scrubbing is in the 90% range. It is very difficult to achieve higher removal efficiencies for this compound. However, HCl is relatively easy to remove from flue gases by scrubbing and an HCl removal efficiency approaching 99.9% may be achieved.

C. HEAT RECOVERY

Heat recovery offers a potential economic advantage by reducing fuel usage with a recuperative heat recovery system or by generating steam with a waste-heat boiler. Combustion temperatures exceeding 1600°F rule out the use of recuperative heat exchangers because of problems with materials of construction and with associated problems such as possible precombustion occurring in the exchangers. A waste-heat boiler can be used effectively with temperatures to $>3000^\circ\text{F}$.

With sulfur-containing VOC in the low heat-content range (<50 Btu/scf) recuperative heat recovery can be used since the combustion temperature does not exceed 1600°F . Waste-heat boilers can be used throughout the operating temperature range for sulfur-containing VOC thermal oxidation. Since halogen-containing VOC destruction always requires high combustion temperatures, only waste-heat boilers can be used for heat recovery. It is possible to use some combination of waste-heat boiler and recuperative-heat recovery, but this option is considered to be beyond the scope of this report.

Special precautions may be needed for the recuperative heaters and waste-heat boilers to prevent excessive corrosion from the corrosive products produced during the combustion of sulfur-containing and halogen-containing VOC. Design considerations should include means of preventing condensation of corrosive gases. For this report the flue-gas temperature after heat recovery will not be cooler than 500°F .

Due to the high heat duties required for waste boilers above 2200°F there will be an upper size limit in terms of waste-gas flow. A waste-gas flow of 50,000 scfm for 2600°F combustion, which would require four large waste boilers, and a waste-gas flow of 20,000 scfm for 3000°F , which would require five large waste boilers, are considered to be the upper size limits in this report. A 200,000,000-Btu/hr heat duty for a waste-heat boiler is considered the upper size limit. Multiple units can be used for higher heat duties.

III. BASIS FOR THERMAL OXIDIZER AND SCRUBBER DESIGN

A base case design based on certain assumptions was developed for this study to generate capital and operating costs, and represents a typical thermal oxidizer installation for the purpose of air pollution control. The effect that varying the design assumptions has on capital and operating costs is discussed. Cost estimates were generated as functions of sensitive design variables, and encompass the accepted range of operation of thermal oxidizers.

A. IDENTIFICATION OF SENSITIVE VARIABLES FOR COST AND ENERGY

A distinction must be made for those design variables which, if changed by a small amount, would cause significant changes in capital cost, annual operating cost, or energy usage. These are called sensitive variables, and the cost curves given in Sect. V include them as parameters. Other variables may be quite important for an individual system design but have relatively minor effects on economic or energy-impact conclusions.

The approach used in this study was to determine the sensitivity of certain variables by means of computerized heat and material balance calculations. Through this process, estimates of the relationships between the variables and equipment design and operating costs may be derived. The primary variables that are a function of the waste gas are the waste-gas temperature, pressure, flow rate, VOC composition (molecular weight, carbon, oxygen, hydrogen, sulfur, and chloride ratios of the VOC, and VOC heats of combustion) and other waste-gas compositions (nitrogen, other inert gases, oxygen, water content, and the presence of special contaminants).

The waste-gas temperature is assumed to be 100°F for the base case, but an increase or decrease within reasonable boundaries will have little effect on the capital or operating costs and it is therefore not considered to be a significant variable. Sensible heat carried by the waste gas is small compared to the heat required to raise the gas to its combustion temperature. About 3.5 Btu/scf is required to increase the temperature of nitrogen from 80°F to 260°F. This compares with fuel heat requirements on the order of 60 to 80 Btu/scf to raise the waste gas to combustion temperatures. Waste-gas temperature differences within this range could not change the total heat requirements

by more than about 6%. Waste-gas pressure is assumed to be 1.5 psig. Pressure changes within reasonable limits of 1.5 psig also will have no significant effect on capital or operating costs. Flow rate is a very significant variable for both capital and operating costs. The waste-gas flows shown in the figures throughout this report are translated to scfm of waste gas to the thermal oxidizer.

The heat content of the waste gas is another significant variable. VOC molar concentration, average molecular weight, carbon, hydrogen, and oxygen ratios, and heats of combustion (Btu/lb of VOC) are all expressed in the variable of the heat content of the waste gas (Btu/scf) as shown by the family of compound lines on Fig. III-1. Multicomponent VOC systems may be described when the mole fractions of each component are known. The contribution of carbon monoxide to the total heat content may also be analyzed in this way. Table III-1 gives VOC molar heats of combustion.⁶

If the actual flue-gas composition is needed, a component material balance must be performed, for which carbon, hydrogen, oxygen, sulfur, and chloride ratios are required. In order to estimate "typical" values for those ratios a group of organic compounds were surveyed.¹ The chlorinated VOC component averages of 34.3% carbon, 4.7% hydrogen, and 61% chlorine were used to establish heat values plus heat and material balance for this evaluation. The sulfur-containing VOC component averages used were the same as the chlorinated-VOC component averages except that 61% was assumed to be sulfur instead of chlorine. The component values of chlorine or sulfur used in this report may be high compared to actual component averages of waste gases, since the VOC in the actual waste gas may not be all chlorinated or sulfur-containing VOC.

The amount of oxygen in the waste gas or that provided by the VOC is important because it establishes the auxiliary combustion air required and has an impact on both the capital and operating costs of the thermal oxidizer. In this report it is assumed that the waste gas and VOC do not contain oxygen (the worst case) and that therefore maximum auxiliary combustion air is required. The excess oxygen in the flue gas is assumed to be 3 mole %, which is based on usually accepted practice.

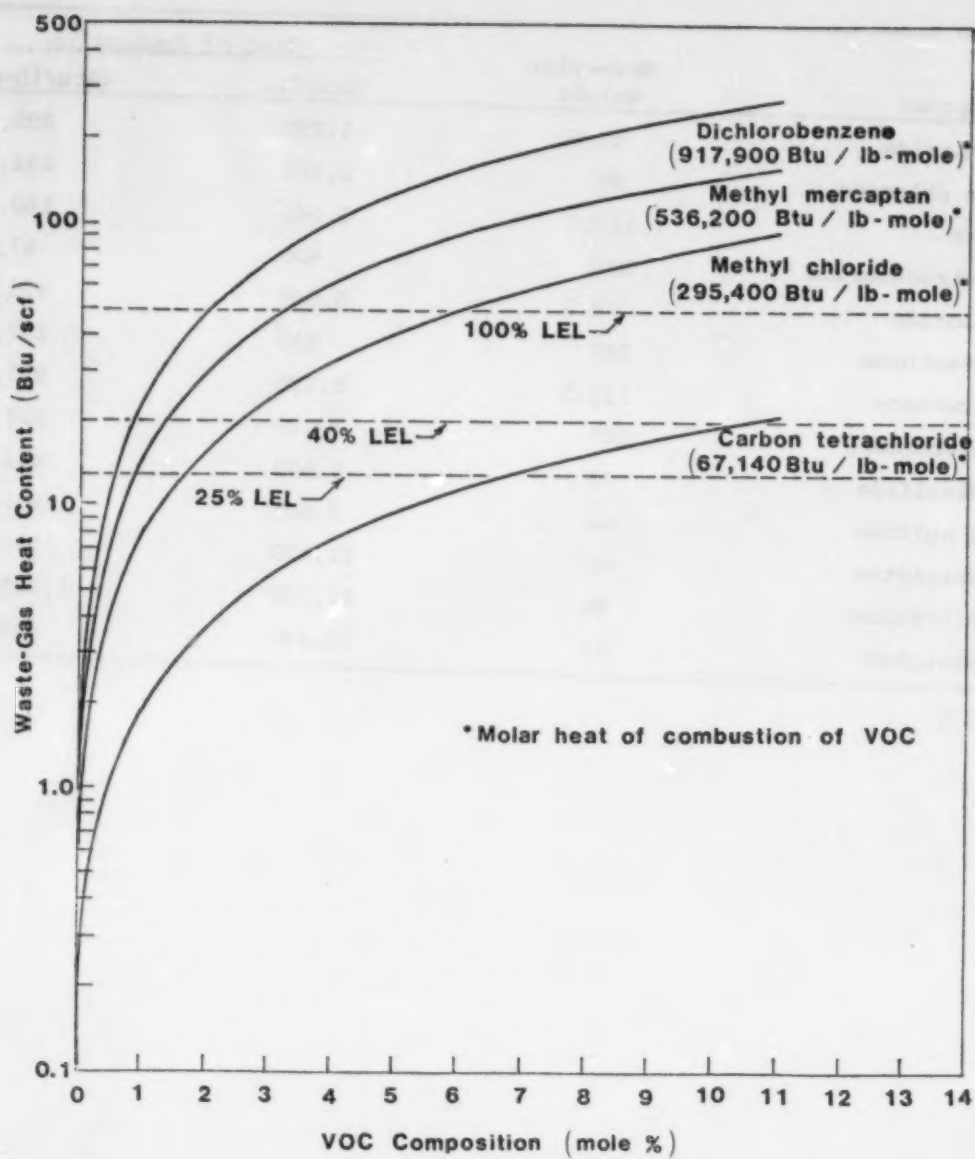


Fig. III-1. Relationship Between Waste-Gas Heat Content, VOC Composition, and VOC Molar Heat of Combustion

Table III-1. VOC Heats of Combustion*

Compound	Molecular Weight	Heat of Combustion	
		(Btu/lb)	(Btu/lb-mole)
Methyl chloride	50.5	5,850	295,400
Methylene chloride	85	2,260	192,100
Chloroform	119.5	1,340	160,100
Carbon tetrachloride	154	436	67,140
Ethyl chloride	64.5	8,840	570,200
Hexachloroethane	237	835	197,900
Dichlorobenzene	111.5	8,230	917,600
Hexachlorobenzene	285	3,220	917,700
Carbon disulfide	76	5,840	443,800
Carbonyl sulfide	60	3,920	235,200
Methyl mercaptan	48	11,170	536,200
Ethyl thiocyanate	96	11,520	1,105,900
Ethyl mercaptan	62	13,130	814,400

*See ref 6.

Since the waste gas is assumed to be saturated with water for calculations in this report, the water content in the flue gas is at a maximum. It is also assumed that no entrained liquid water droplets enter with the feed. Entrained water droplets would significantly increase the auxiliary fuel requirements since the heat capacity of the flue gas increases and the heat of vaporization for water, 18,000 Btu/lb-mole, must be supplied. Entrainment can normally be avoided with proper design.

The combustion temperature has a significant effect on fuel costs. As is shown in Fig. II-2, the fuel usages increase substantially with temperature increases above 2200°F. Residence time has a major effect on capital costs but a relatively small effect on operating costs.

The significant variables investigated in this report are waste-gas flow, heat content (encompassing a variety of composition-related variables), combustion temperature, and residence times. Evaluation of these parameters using generally worst-case or conservative assumptions will lead to economic and energy impact conclusions that will equal or exceed actual operating costs and energy impacts for the applicable cases.

B. PROCEDURE USED FOR DESIGNING A THERMAL OXIDIZER SYSTEM WITH A SCRUBBER

The design procedure that was used for conventional thermal oxidizer units for sulfur-containing VOC is the same as that given in the control-device evaluation report for thermal oxidation, except for the scrubbers. The discussion of quench chambers and scrubbers in this section applies to conventional thermal oxidizers that may require the addition of flue-gas scrubbing. The design procedure for high-temperature thermal oxidation is presented in this section.

1. Combustion Chamber

The size of the combustion chamber depends on the flow of waste gas into the chamber, the fuel usage, the residence time, and the combustion temperature. The fuel usage is dependent on the heat content of the waste gas and the combustion temperature. Figure III-2 shows the supplementary fuel usage of natural gas required to sustain the combustion temperatures of 1800, 2200, 2600, and 3000°F for various waste-gas heat contents. The combustion chamber, the heat recovery

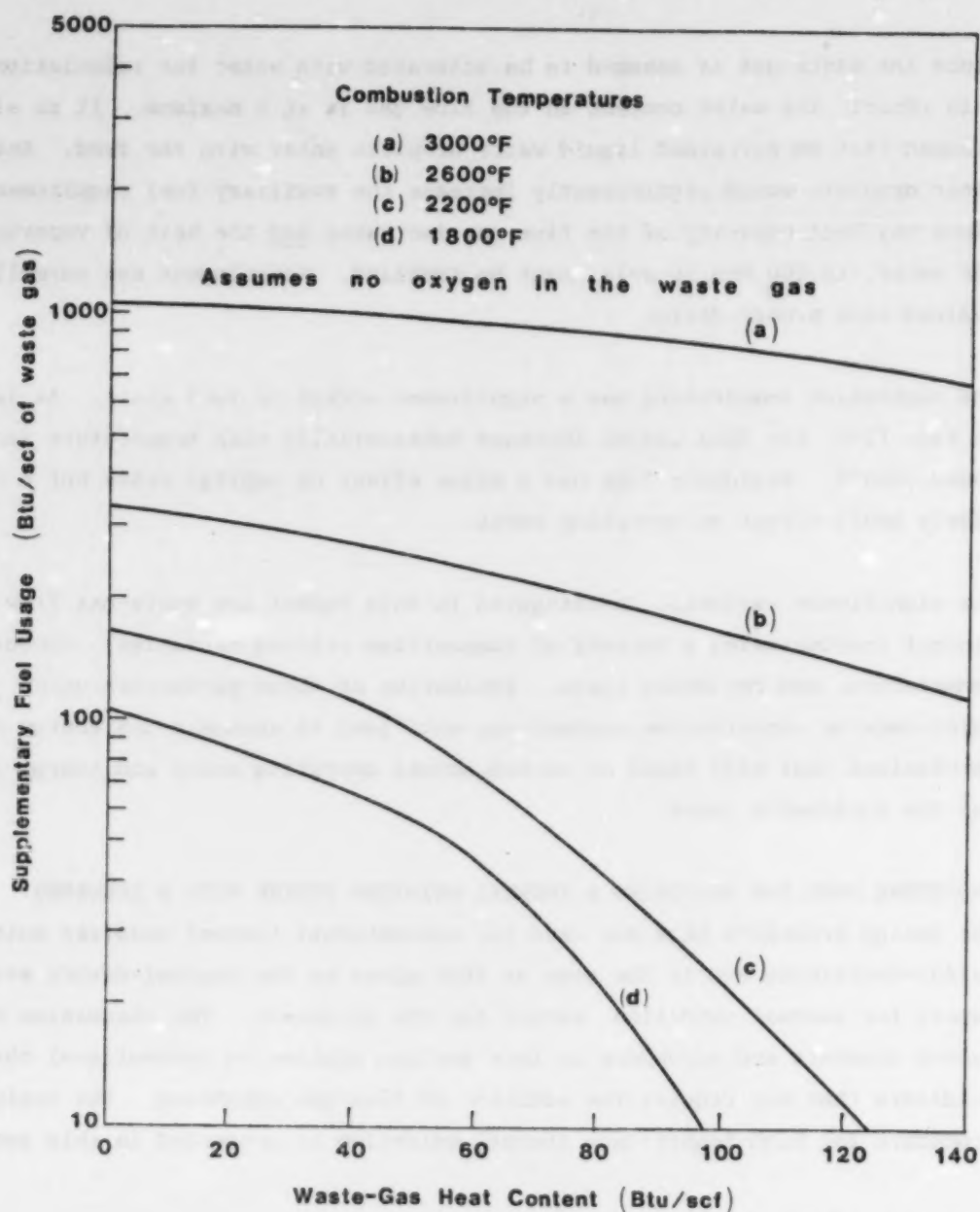


Fig. III-2. Supplementary Fuel Usage vs Waste-Gas Heat Content

unit, and the scrubber must be sized to handle the flow from all sources, including that from the waste gas, the auxiliary fuel, the combustion air, and the combustion products. System designs calculated for this report cover the waste-gas heat content range of 1—100 Btu/scf. Figure III-3 shows the correlation between flue-gas flow and waste-gas flow by expressing them as a ratio and displaying the ratio as a function of waste-gas heat content and combustion temperature. In Fig. III-3 the 1400 and 1600°F lines curve upward above 80 Btu/scf due to the need to add excess air to keep the combustion temperature of 1400 and 1600°F. The temperature would normally be allowed to increase rather than excess air being added.

Conversion of scfm (standard cubic feet per minute) to acfm (actual cubic feet per minute) is necessary for sizing of the combustion chamber. Figure III-4 shows this relationship (based on the ideal gas law). Standard temperature and pressure conditions assumed throughout this report are 32°F and 760 mm Hg and their equivalents. The ratio of acfm to scfm is read from Fig. III-4 and is multiplied by the combustion chamber flue gas in scfm. Until this point in the report the term heat content has referred to the potential energy from combustion of compounds in the waste gas. Heat content when used in relation to the flue gas in this report refers to the energy contained by the hot gases because of their temperature. The basis used for the heat recovery calculations is the heat content of the flue gas as a function of temperature, as shown in Fig. III-5. The dotted lines in Fig. III-5 correspond to a reasonable variation in specific heats or heat capacities of the flue gas.

Residence times of 1/2 and 3/4 sec are used to calculate the combustion chamber internal volume. The combustion chamber flue gas flow (in acfm) is converted to acfs (actual cubic feet per second) and then multiplied by the residence time (in sec) to determine the combustion chamber internal volume in cubic feet.

2. Fans

Fans for both the waste gas and the combustion air are provided for in the systems evaluated. The flow rates of the waste gas and combustion air and the pressure drops of the thermal oxidizers are used to calculate fan sizes. System pressure drops of 6 and 10 in. H₂O were assumed respectively for thermal

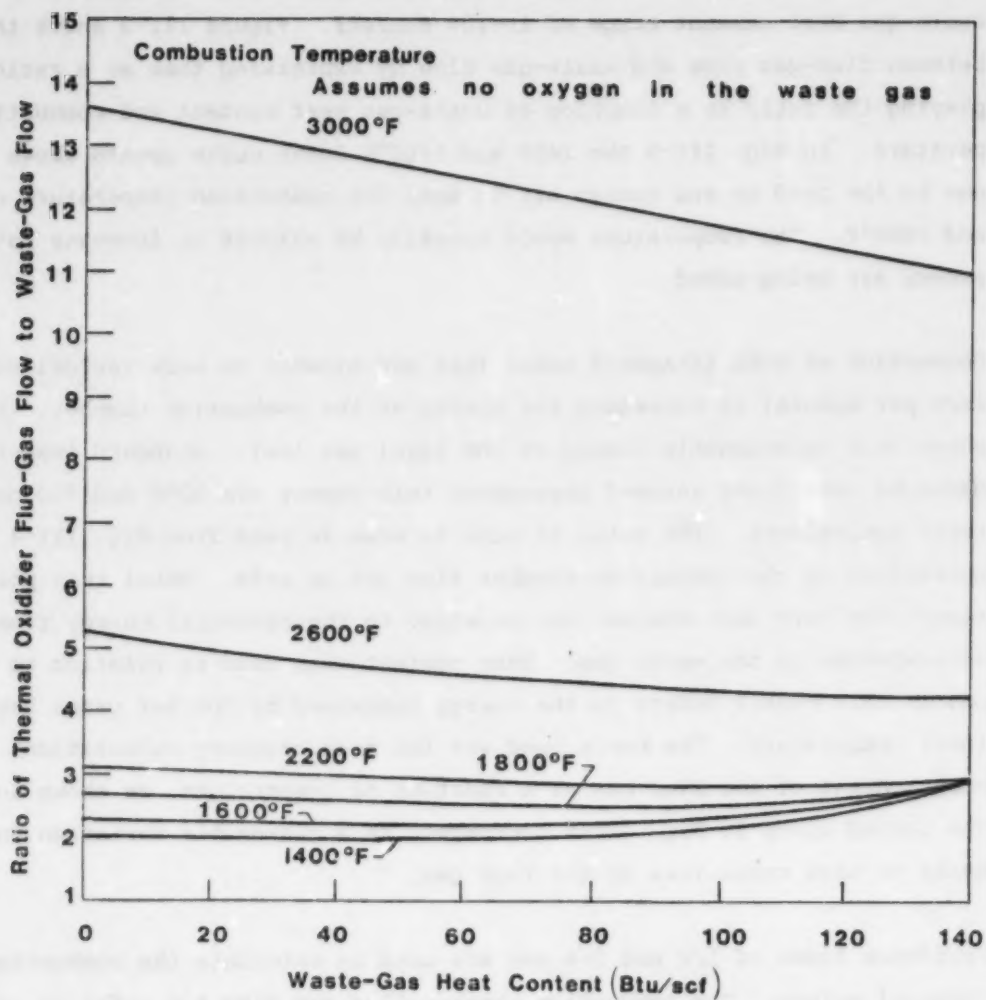


Fig. III-3. Ratio of Thermal Oxidizer Flue-Gas Flow to Waste-Gas Flow vs Waste-Gas Heat Content

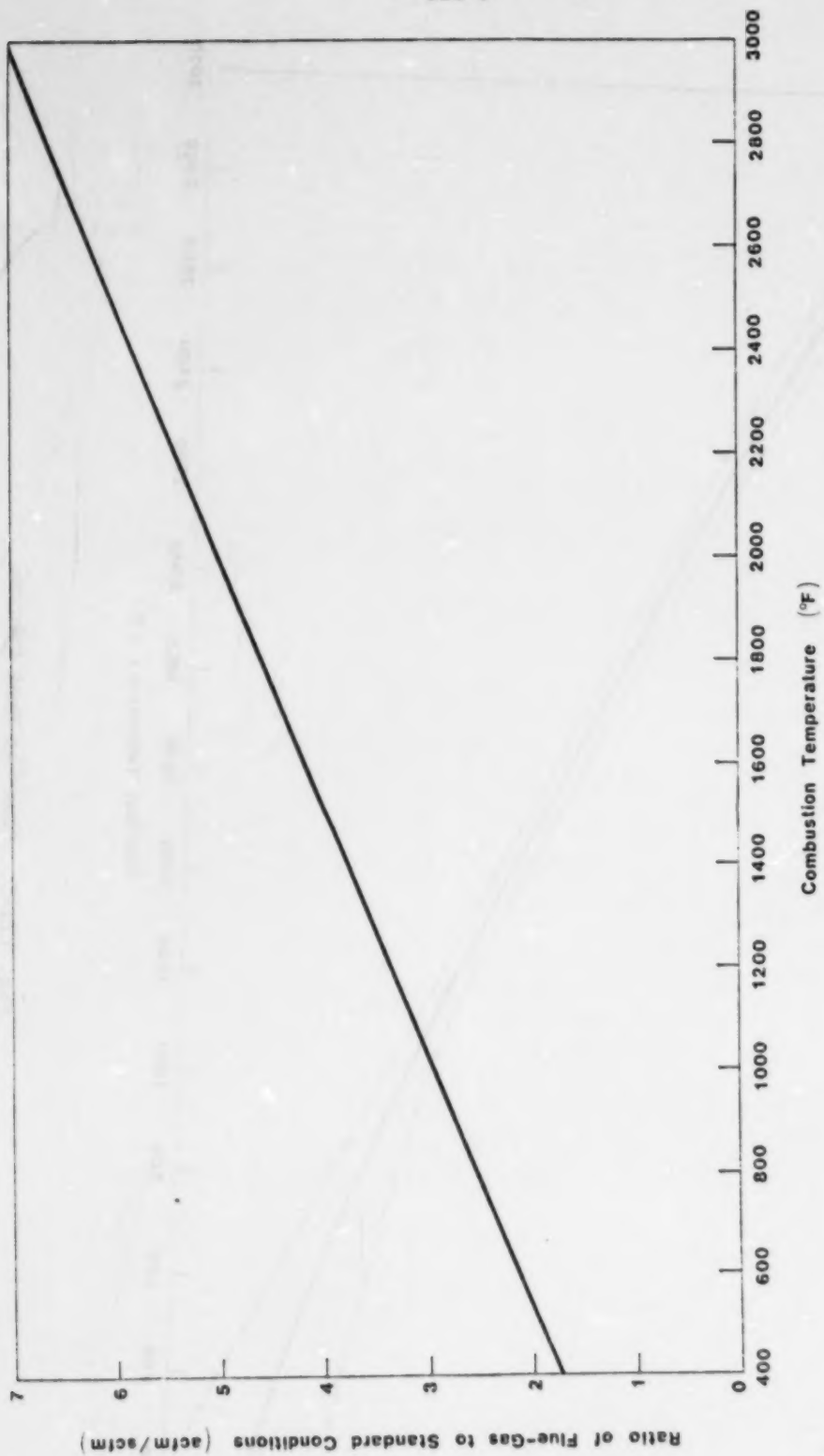


Fig. III-4. Relationship Between Actual Flow Rates and Flue-Gas Temperature

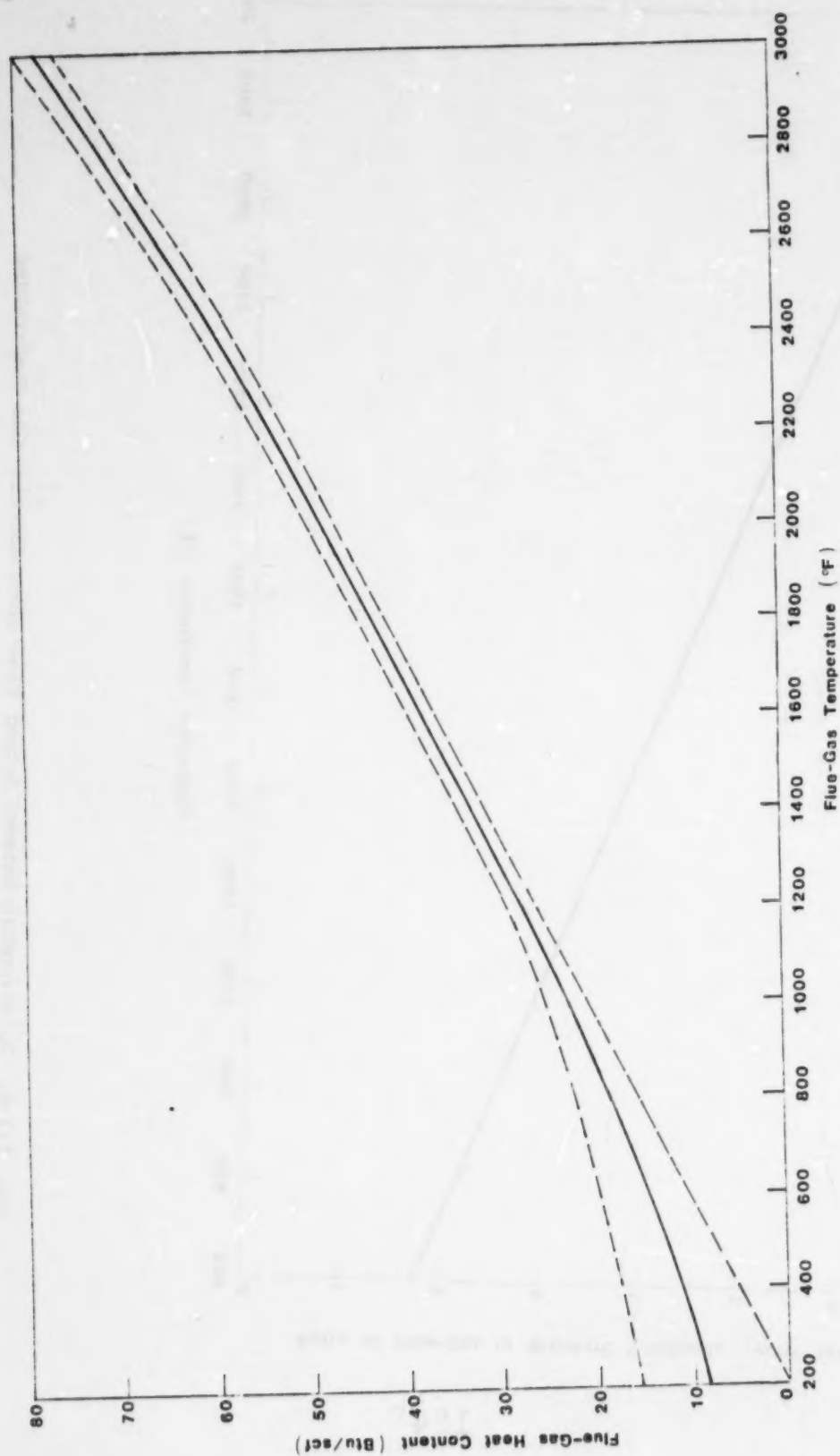


Fig. III-5. Flue-Gas Heat Content

oxidizers with no heat recovery and those with waste-heat boilers. A pressure drop of 12 in. H_2O is assumed when a scrubber is included. The waste-gas fan capacity is based on the waste-gas flow rate. Table III-2 gives the combustion air flow/waste-gas ratio as a function of combustion temperature. The relationship is based on desired combustion temperatures, as previously discussed, and is used to size the combustion air fan. The combustion air volume and fan size may be determined by the use of this ratio multiplied by the volume of waste gas. Waste gases containing significant levels of oxygen would reduce the combustion air required and the size of the combustion air fans used in this study. When waste gases are generated at higher pressure, often no fan is needed.

3. Waste Heat Steam-Generation Boiler Heat Recovery

The heat recovery by a waste-heat boiler depends on the entrance flue-gas temperature and the exhaust temperature. The entrance temperature is set by the thermal oxidizer conditions. For this study the exhaust temperature is assumed to be 500°F, in order to be well above the flue-gas dew point to prevent potential corrosion from acid gases. Figure III-6 is a plot of the maximum heat and is based on the percent of heat available from the flue gas with standard conditions as the reference point. The dotted lines in the figure relate to a reasonable range of heat capacities or specific heats of the flue gas. Figure III-7 relates the ratio of boiler-tube surface area to the flue-gas flow rate, the flue-gas temperature, the steam temperature, and the overall heat transfer coefficient. The surface area of the waste-heat boiler is determined by multiplying the flue-gas flow rate by the ratio from Fig. III-7. The flue-gas flow rate from a thermal oxidizer employing a waste-heat boiler is the same as that from a thermal oxidizer using no heat recovery.

4. Quench Chamber

The quench chamber in this study is located in the lower part of the scrubber column and has the same diameter as the scrubber. The volume is based on a 1-sec flue-gas retention time. The quench reduces the flue-gas temperature to the adiabatic saturation temperature of the scrubber agent. For this design slightly alkaline water will be used as the scrubbing agent. To reduce the flue-gas temperature to the adiabatic saturation temperature of water, considerable

Table III-2. Ratio of Combustion Air to Waste-Gas Flow Rate vs Combustion Temperature

Waste-Gas Heat Content (Btu/scf)	Combustion Temperature (°F)	Combustion Air to Waste-Gas Flow Ratio* (scf of combustion air / scf of waste gas)
2	1400	0.87
2	1600	1.1
50	1800	1.2
100	2200	1.4
100	2600	3.1
100	3000	9.8

*Thermal oxidizer conditions:

No oxygen in waste gas.

VOC molar heat of combustion = 746,000 Btu/lb-mole.

VOC molecular weight = 100

VOC C, H, Cl fraction = 34.3 wt % C, 4.7 wt % H, 61 wt % Cl.

Average waste gas molecular weight = 29.

Water content of combustion air = 1.0 wt %.

3 mole % O₂ in flue gas after oxidation.

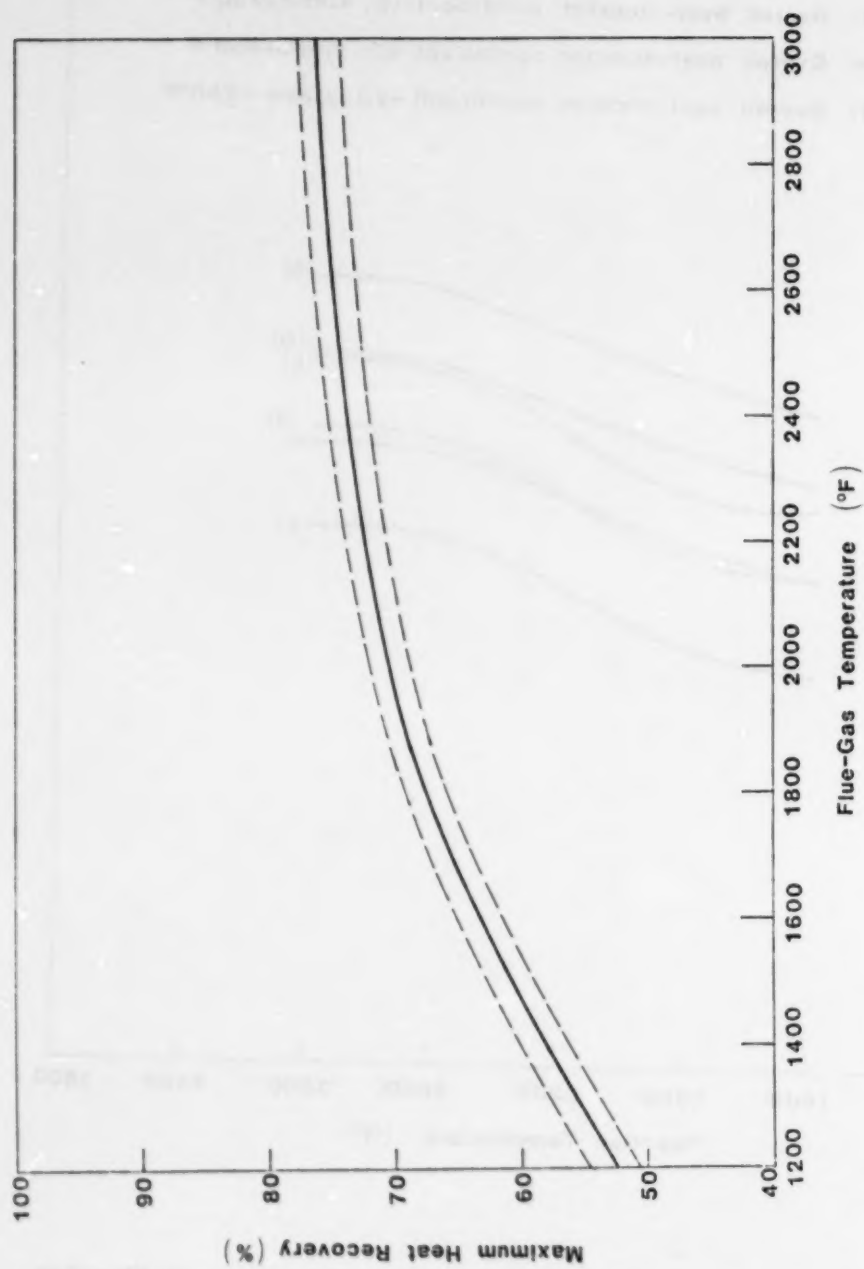


Fig. III-6. Maximum Heat Recovery from a Waste-Heat Boiler
(Exhaust Temperature = 500°F)

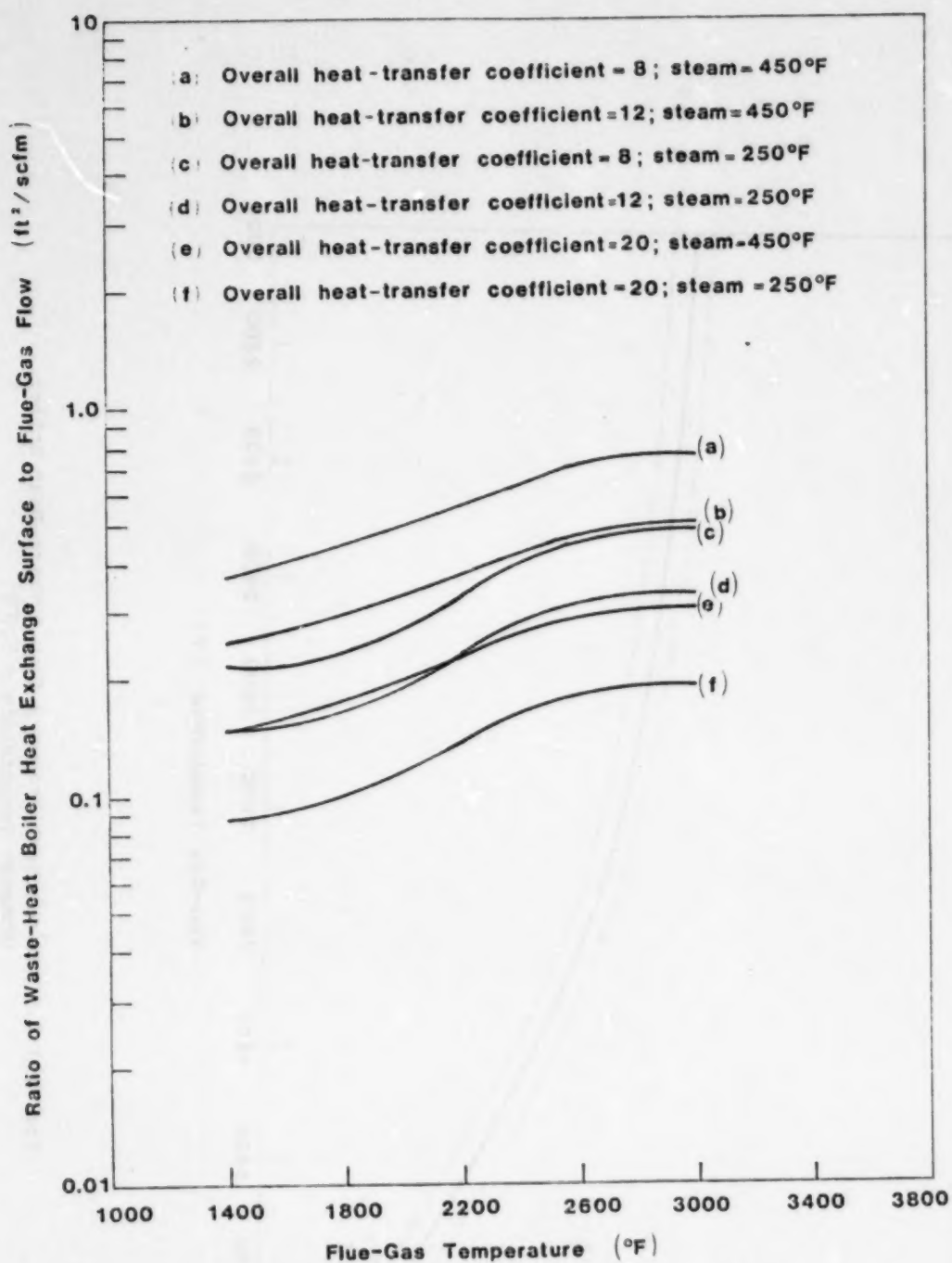


Fig. III-7. Ratio of Waste-Heat Boiler Heat Exchange Surface to Flue-Gas Flow vs Flue-Gas Temperature

water will be vaporized in the quench and will increase the total gas flow through the scrubber. This increased flow rate through the scrubber due to the vaporization of water, which is a function of temperature, is shown in Fig. III-8.

The quench ratio has to be multiplied by the flue-gas rate to obtain the flow rate through the scrubber. Makeup water usage for the quench is dependent on the flue-gas flow rate and temperature. The makeup water rate is shown in Fig. III-9. The quench will remove some of the noxious gases in the flue gas. In this study it has been assumed that 50% of HCl is removed in the quench and that 10% of SO₂ is removed in the quench.

5. Scrubber

The scrubber design in this study is based on a packed column, with slightly alkaline water used as the scrubbing agent. Different designs for the contacting device and its aqueous scrubbing agent could be used, but they would not significantly offset the capital cost presented. Operating cost could be significantly different based on the alkaline agent used to neutralize the scrubbed acid gases. Acid gas removal efficiencies would also be affected by the choice of designs.

The column is designed with 36 ft of packing, which is assumed to remove 99.8% of HCl or 88.9% of SO₂ in the scrubber. These two removal percentages combined with the 50% of HCl or 10% of SO₂ removed in the quench give a total removal efficiency of 99.9% HCl or 90% SO₂. The liquid-to-gas ratio (L/G) is assumed to be 10. A superficial vapor velocity of 3 fps was used for determining the column diameter. These assumptions are adequate for initial process designs leading to a preliminary cost estimate.⁷

The water used as the scrubbing agent will have to be neutralized to control the pH of the system; in this study caustic (NaOH) is used. The amount of caustic used is dependent on the concentration of sulfur or chlorine in the waste gas. The caustic addition rate is 2.50 lb of NaOH/lb of sulfur in the waste gas and 1.14 lb of NaOH/lb of chlorine in the waste gas. The salt formed will have to be purged from the system and discharged. The makeup water rate, based on 1% dissolved solids in the water recycle, is 46.5 gal/lb of sulfur in the waste gas and 19.2 gal/lb of chlorine in the waste gas .

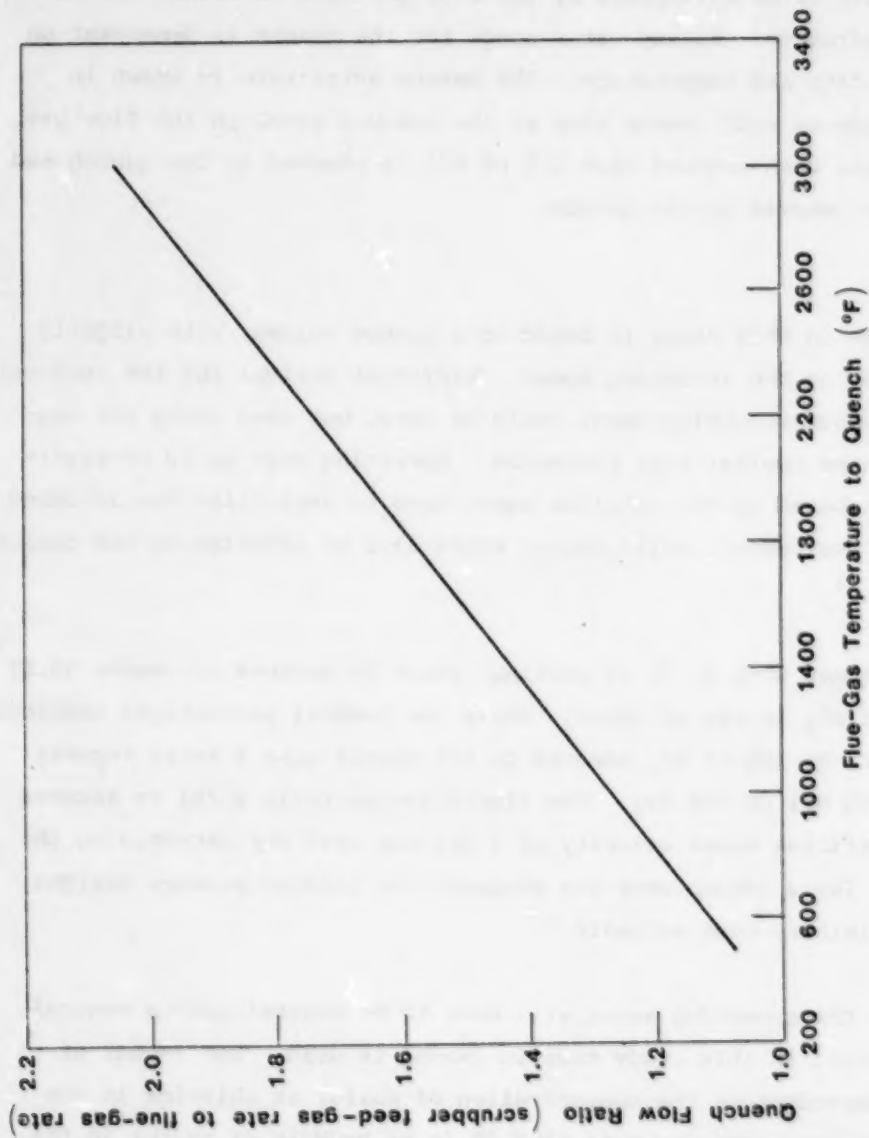


Fig. III-8. Ratio of Quenched to Unquenched Flue-Gas Flow Rate vs Flue-Gas Temperature

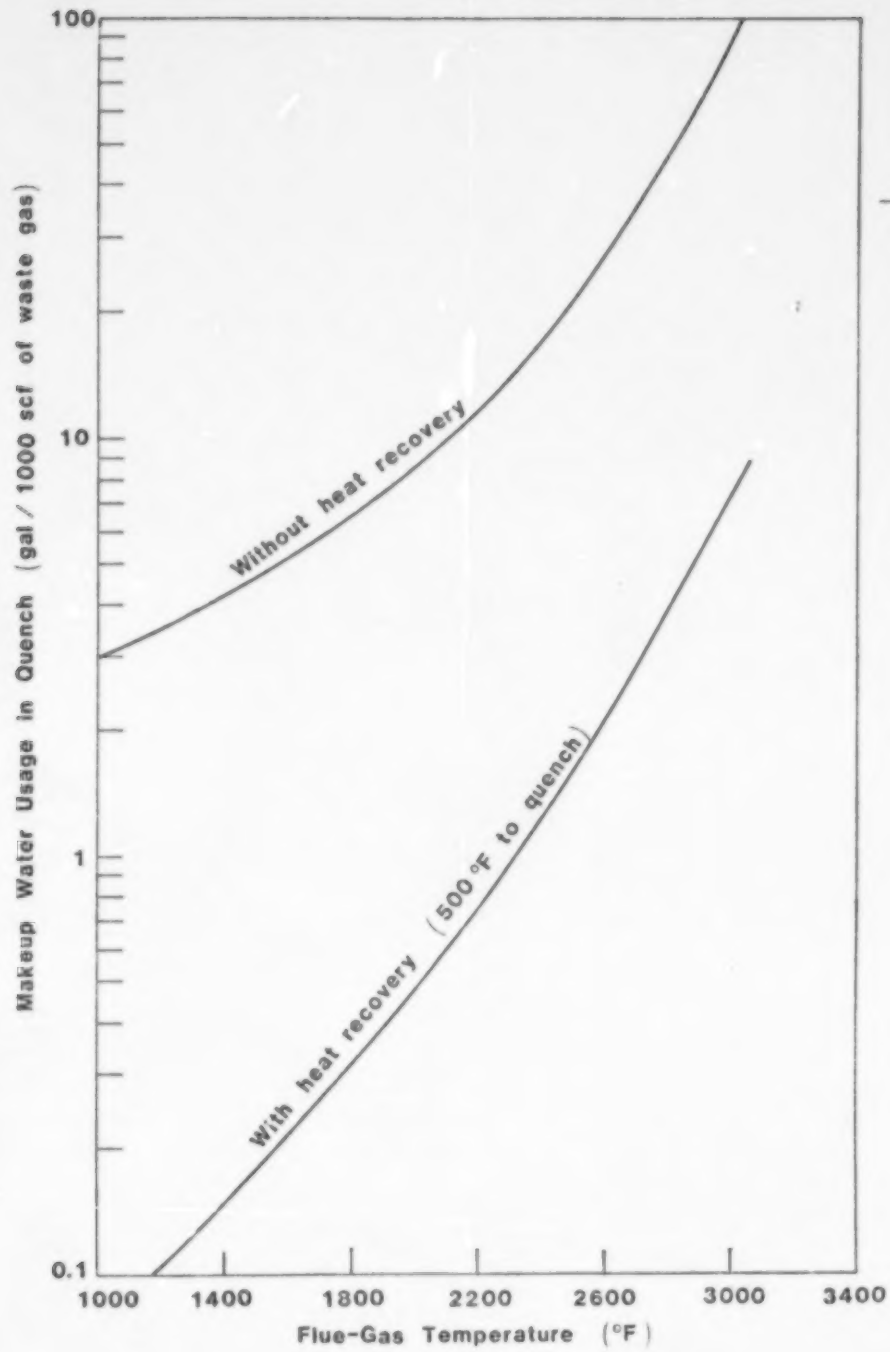


Fig. III-9. Makeup Water Usage Rate for Quenching Chamber

IV. CONSIDERATIONS FOR INSTALLING THERMAL OXIDATION CONTROL EQUIPMENT

Thermal oxidizers can be large process units, depending on the volume of waste gas to be controlled, and could require a plot of land as large as 300 ft by 300 ft for installation. Since thermal oxidizers utilize combustion with a flame for achieving VOC destruction, the unit must be located at a safe distance from process equipment in which flammable chemicals are used, or special designs must be employed to minimize the risk of explosion or fire.

Thermal oxidizers require natural gas or fuel oil, electrical power, and instrument air and, if scrubbing is needed, water and caustic at the site. If steam is generated from waste heat, then it is useful to minimize the distance from the waste-heat boiler to the steam-consuming site.

Since a salt is formed during the scrubbing of the flue gas, proper waste disposal of this material is required. The options of disposal range from direct wastewater discharge to recovery of the material.

Retrofitting thermal oxidizers into existing plants requires careful consideration of site location since all the above factors apply and sufficient space in an existing plant may not be available. The unit may have to be located further away from the waste-gas source than would be required for a new plant. Because of these associated costs the cost of retrofitting a thermal oxidizer in an existing plant may be appreciably greater than the cost for a new installation. Also, since it may be costly for some companies to have excess steam on-site, it may not be practical for all companies to utilize the heat recovery option.

V. COST AND ENERGY IMPACTS OF THERMAL OXIDIZERS WITH SCRUBBERS

A. COST BASIS

The capital costs for total systems combinations and for various components were estimated. They represent the total investment, including all indirect costs such as engineering and contractors' fees and overheads, required for purchase and installation of all equipment and material to provide a facility as described. These are battery-limit costs and do not include the provisions for bringing utilities, services, or roads to the site, the backup facilities, the land, the research and development required, or the process piping and instrumentation interconnections that may be required within the process generating the waste-gas feed to the thermal oxidizer.

The estimated costs are based on installation of a new plant; no retrofit cost considerations are included. Those costs are usually higher than the cost for a new-site installation for the same system and include, for example, demolition, crowded construction working conditions, scheduling construction activities with production activities, and longer interconnecting piping. Since the thermal oxidizer systems require a relatively large land area and the safety aspects of an open flame are an important factor, the longer interconnecting piping will probably be the most significant of these retrofit cost factors. These factors are so site-specific that no attempt has been made to provide costs. For specific retrofit cases rough costs can be obtained by using the new-site data and adding as required for a defined specific retrofit situation.

The method used to develop these estimated capital costs was based on preliminary vendor quotes for the purchase of major equipment items or from such sources as Richardson Engineering Co. data, and factoring up to installed costs based on the data in Table V-1. The expected accuracy of the total installed cost with this degree of engineering detail using this factor method is $\pm 30\%$. This method of obtaining estimated total installed capital costs is suitable for a cost study or for screening estimates. The factor ranges given in Table V-1 for various cost components are based on historical data obtained by Hydrosience Process Engineering.

Table V-1. Factors Used for Estimating Total Installed Costs

A = Major Equipment Purchase Cost Plus 0.1 to 0.35 Allowance

Installation Costs

Foundations	0.06A + \$100 X number of pumps
Structures	0.15A (no structures) to 0.30A (multideck structures)
Equipment Erection	0.15A to 0.30A (depending on complexity)
Piping	0.40A (package units) to 1.10A (rat's nest)
Insulation	0.06A or 0.15 X piping (normal) to 0.30 X piping (bulk hot or cold)
Paint	0.05A
Fire Protection	0.01A to 0.06A (depending on requirements)
Instruments	0.10A to 0.30A or 0.01A to 0.25A + \$50,000 to \$300,000 for process control computer
Electrical	0.15A or 0.05A + \$500 per motor

B = Base Cost A + Sum of Installation Costs

Sales Tax 0.025A + 0.025B

Freight 0.16A

Contractor's Fees 0.30 (B-A)

C = Total Contract B + Taxes, Freight, and Fees

Engineering^a 0.01C to 0.20C

Contingencies^b 0.15C

D = Process Unit Installed Cost C + Engineering + Contingencies

E = Total Subestimates

Sum of semidetalled subestimates (buildings, site development, cooling towers, etc.). Each subestimate should include taxes, freight, fees, engineering and contingency, and should be escalated to date of expenditure for that cost component. Engineering costs, contingencies, and escalation factors for these subestimates will vary according to the type of job.

F = Total Project Cost

D + E

^a Includes cost from capital project teams, process engineering, engineering, purchasing, and other support groups.

^b Contingency should not be applied to any cost component that has been committed by either purchase order or contract.

The estimate is based on the purchase cost of major equipment (A), including a 10 to 35% allowance for other equipment and an assessment of the quality of vendor quotes. A 10% allowance is used for project definition that includes process flow sheets and specific budget quotes and a 35% allowance for block flow sheet definition and generalized equipment quotes or prices.

B. CAPITAL COSTS

1. Thermal Oxidizer Complete System

The capital cost for thermal oxidizer systems controlling sulfur-containing VOC is determined from the capital costs estimated in the control-device evaluation report for thermal oxidation¹ (Figs. V-1 to 4). The installed capital cost of the scrubber shown in Fig. V-1 can be added to the installed capital cost given in the cited report¹ to obtain the total installed capital cost of the system.

The capital cost for thermal oxidizer systems controlling halogen-containing VOC is determined by combining the capital cost of the thermal oxidizer (Figs. V-2, 3), of the waste-heat boiler (Figs. V-4, 5), and of the scrubber (Fig. V-1). At the combustion temperature of 1800 and 2200°F the capital costs are obtained from the control-device evaluation report for thermal oxidation.¹ The total installed system costs are shown in Figs. V-6 through V-11. Figures V-9 through V-11 are based on a waste-gas heat content of 1 Btu/scf. An increase in waste-gas heat content will decrease the flue-gas flow rate, as shown in Fig. III-3, and decrease the scrubber size.

2. High-Temperature Thermal Oxidizers

Preliminary purchase quotes for high-temperature thermal oxidizers were obtained from two vendors.^{8,9} The installed costs shown in Fig. V-3 include the cost of fans, ductwork, and stacks. The upper size limit for the high-temperature thermal oxidizers is assumed to be 20,000 cfm for a 2600°F combustion temperature and to be 5000 cfm for a 3000°F combustion temperature. The reason for this limit is due to the waste stream not containing oxygen so large amounts of combustion air is required.

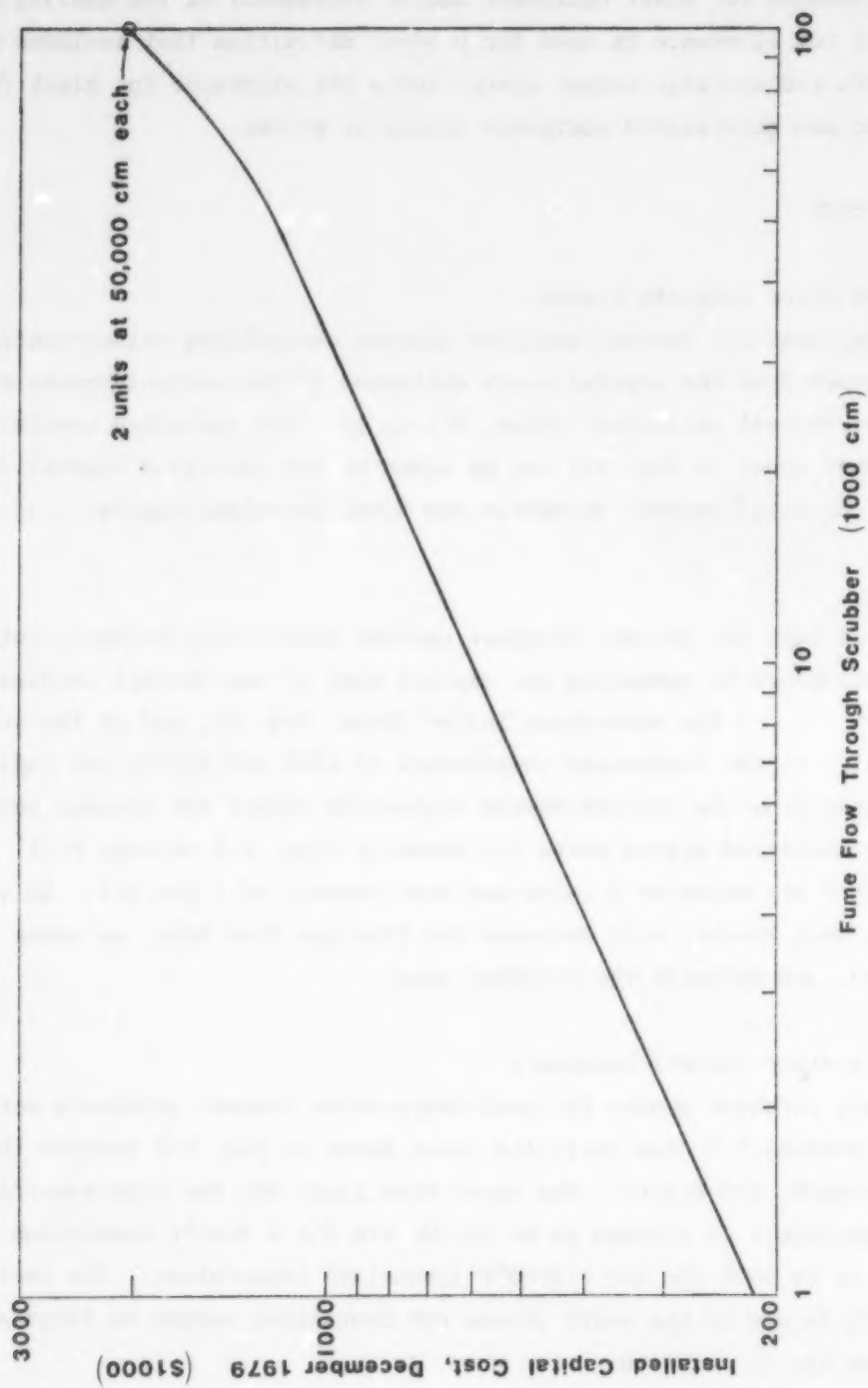


Fig. V-1. Installed Capital Cost of the Scrubber Including Quench Chamber

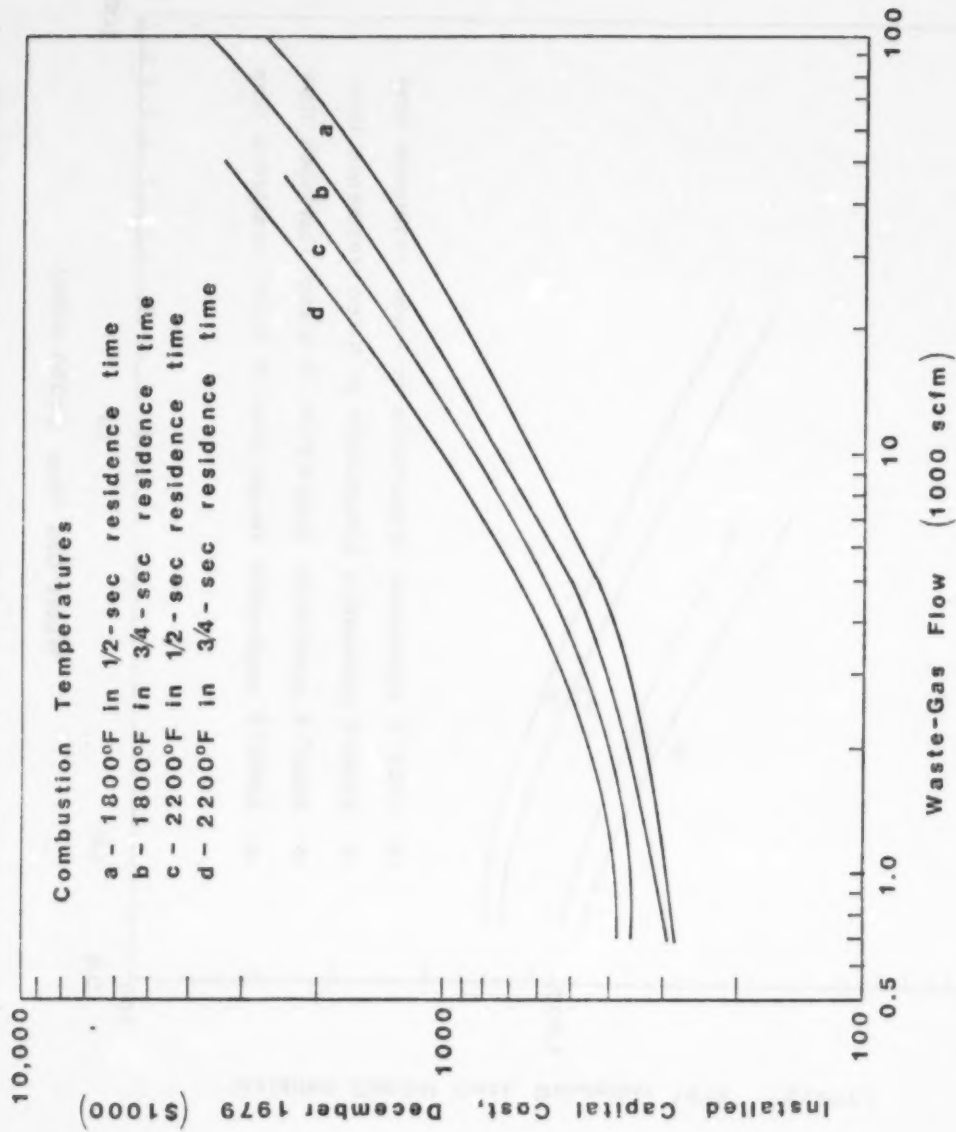


Fig. V-2. Installed Capital Cost of Thermal Oxidizer at 1800 and 2200°F Including Incinerator, Two Blowers, Ducts, and Stack

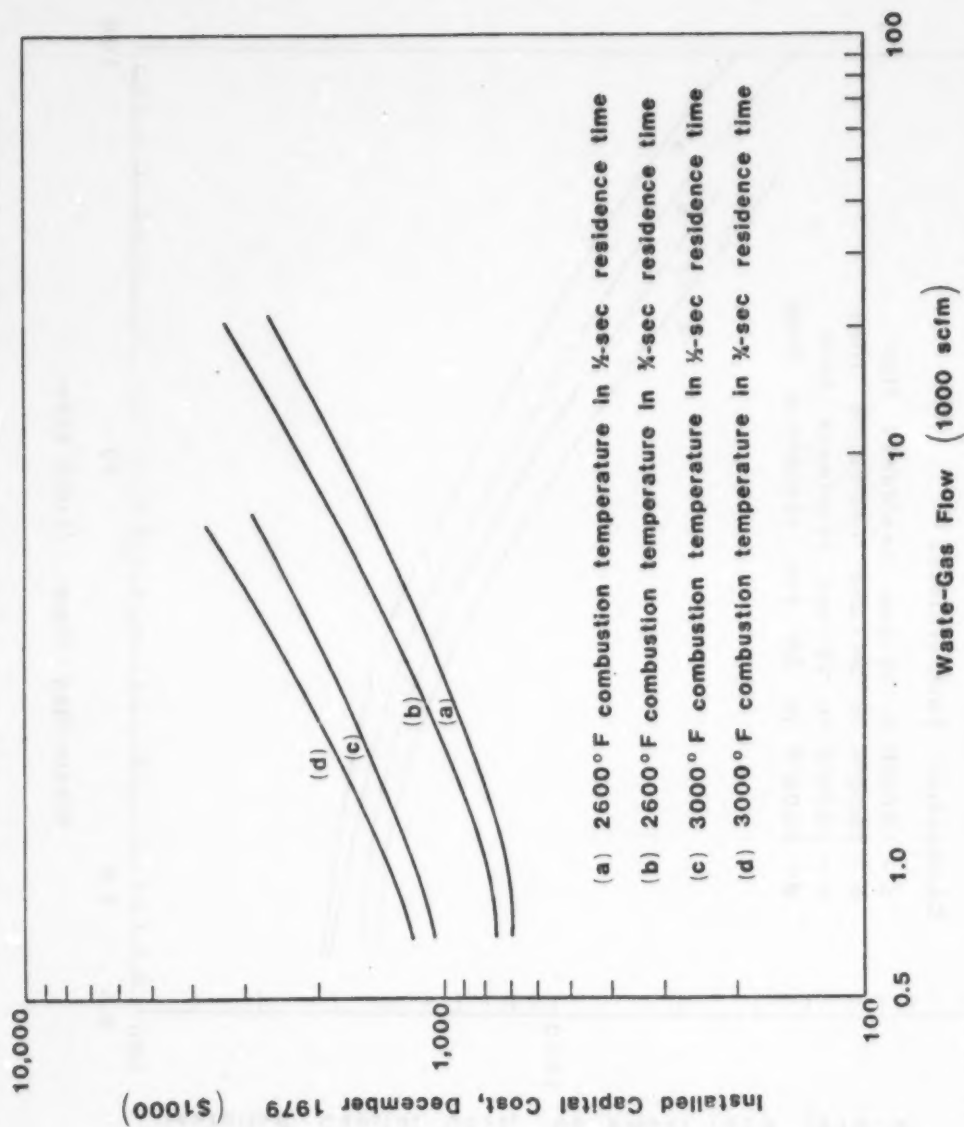


Fig. V-3. Installed Capital Cost of Thermal Oxidizer at 2600 and 3000°F, Including Incinerator, Two Blowers, Ducts, and Stack

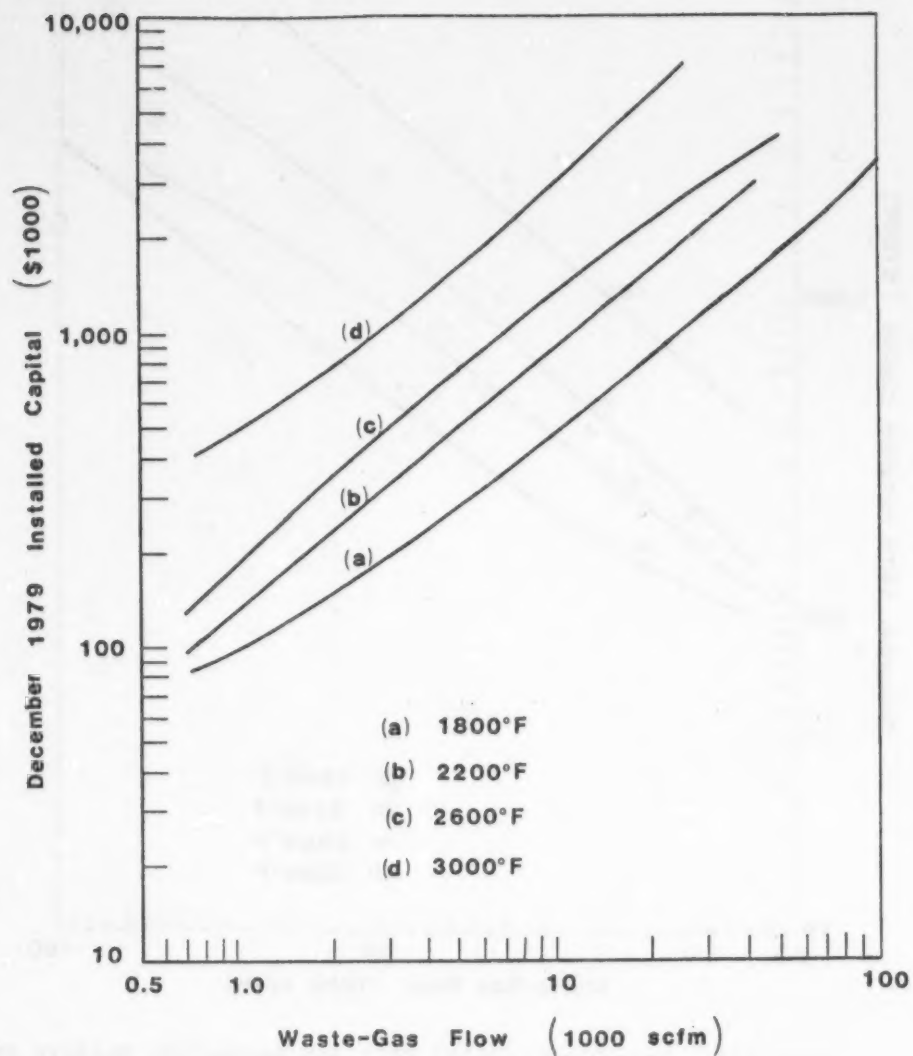


Fig. V-4. Installed Capital Cost for Waste-Heat Boilers at 250 psi and 1800, 2200, 2600, and 3000°F Combustion Temperatures

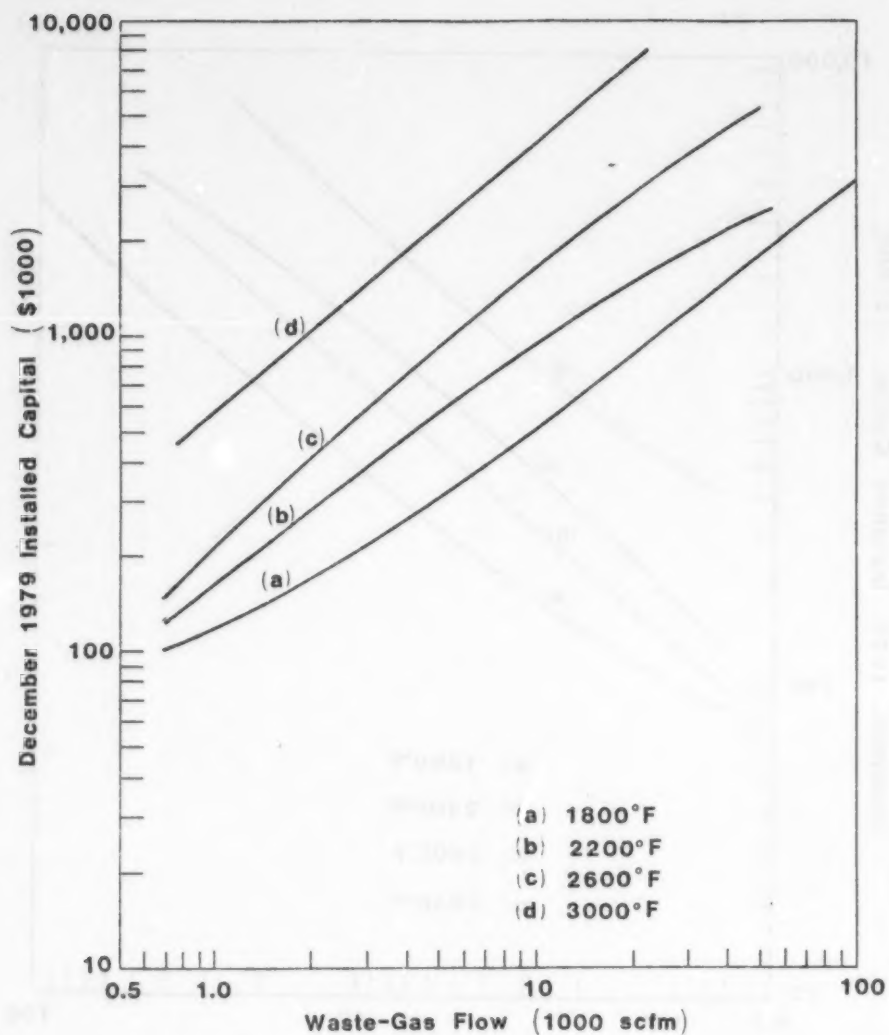


Fig. V-5. Installed Capital Cost for Waste-Heat Boilers at 400 psi and 1800, 2200, 2600, and 3000°F Combustion Temperatures

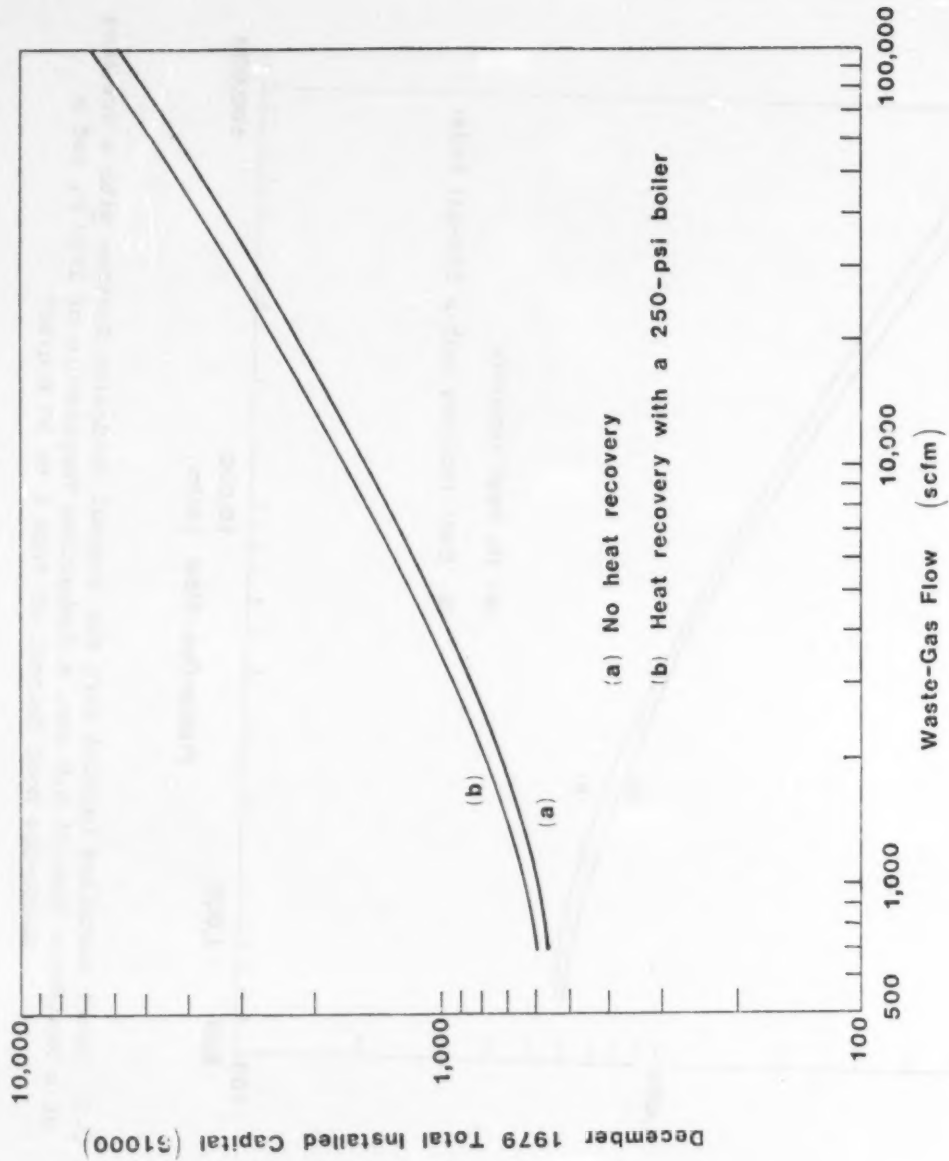


Fig. V-6. Total Installed Capital Cost for Thermal Oxidation Systems with a Scrubber at a Residence Time of 0.5 sec, a Combustion Temperature of 1400°F, and a Waste-Gas Heat Content from 1 to 50 Btu/scf

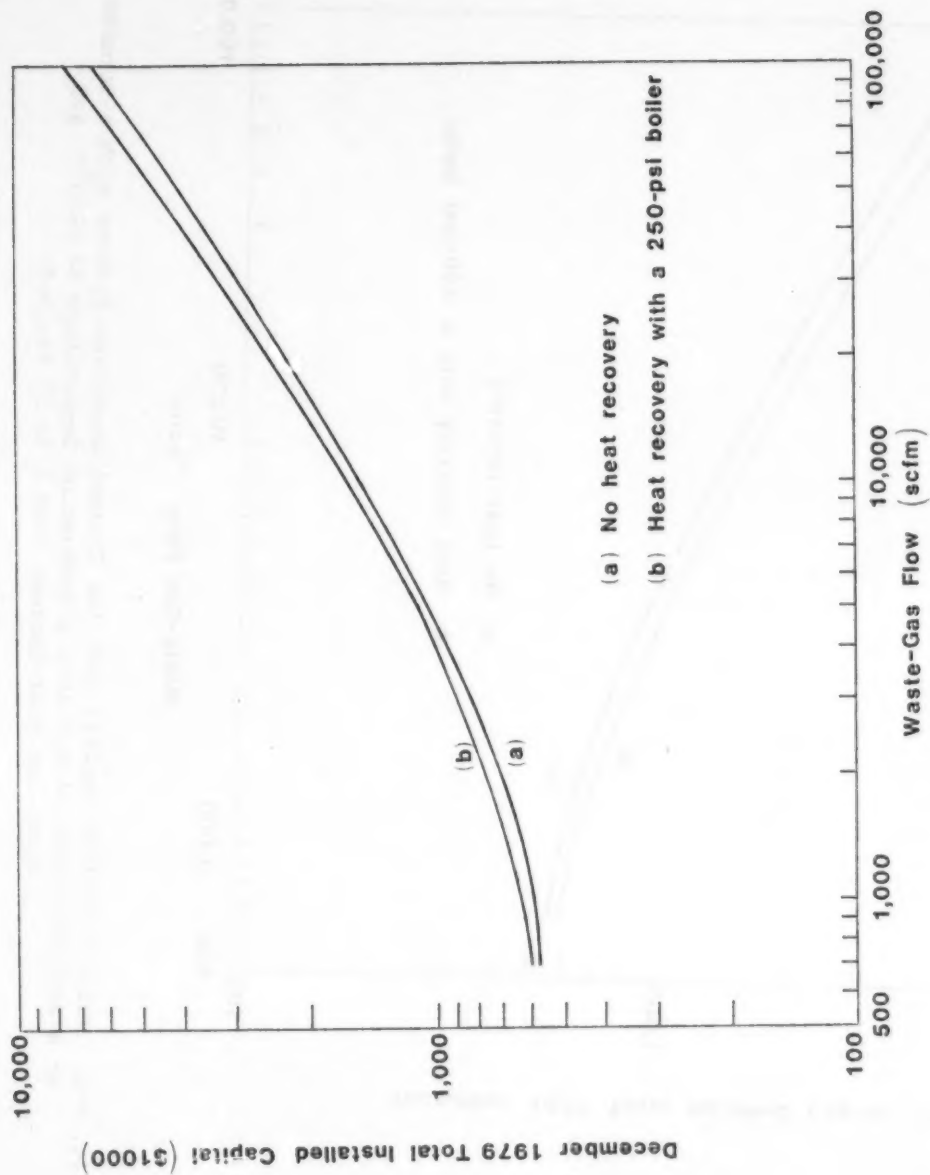


Fig. V-7. Total Installed Capital Cost for Thermal Oxidation Systems with a Scrubber at a Residence Time of 0.5 sec, a Combustion Temperature of 1600°F, and a Waste-Gas Heat Content of from 1 to 50 Btu/scf

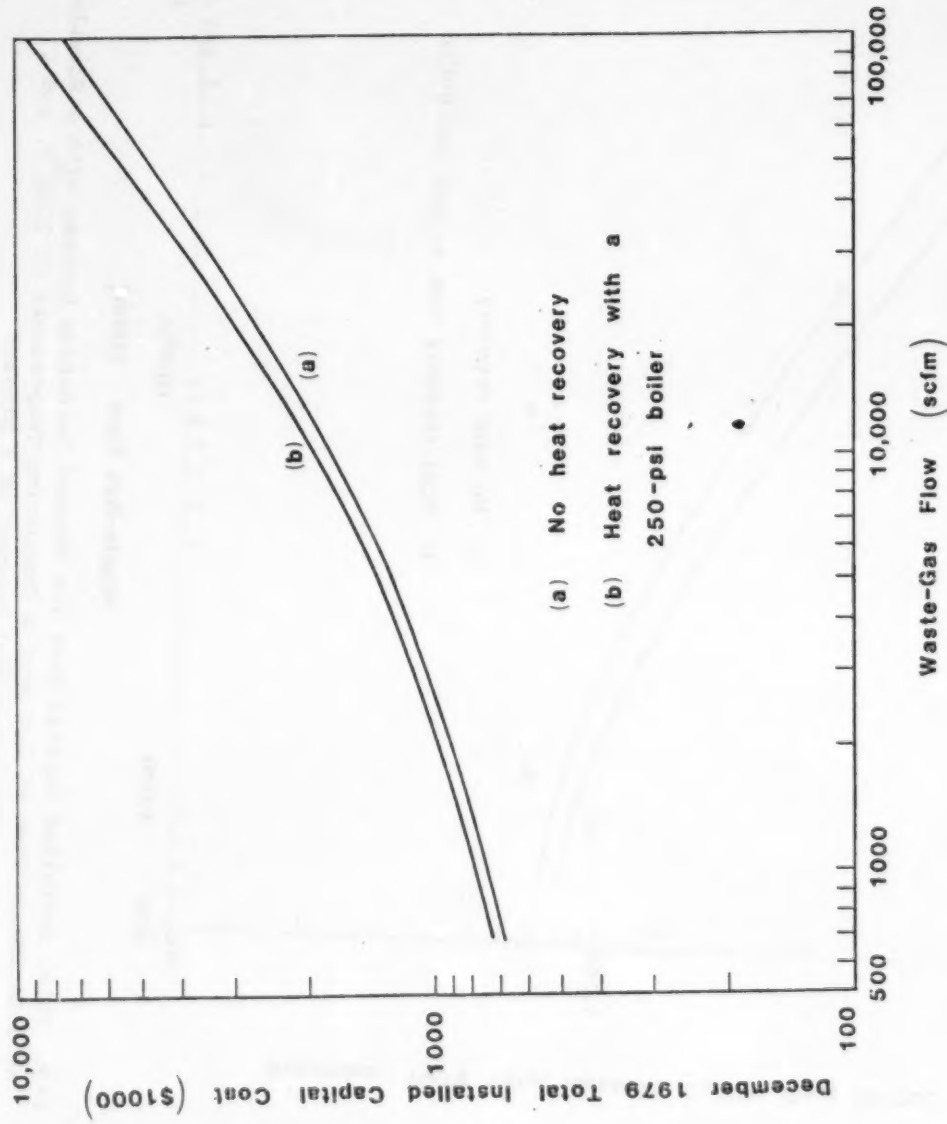


Fig. V-8. Total Installed Capital Cost for Thermal Oxidation Systems with a Scrubber, at a Residence Time of 0.5 sec, a Combustion Temperature of 1800°F and a Waste-Gas Heat Content of from 1 to 50 Btu/scf

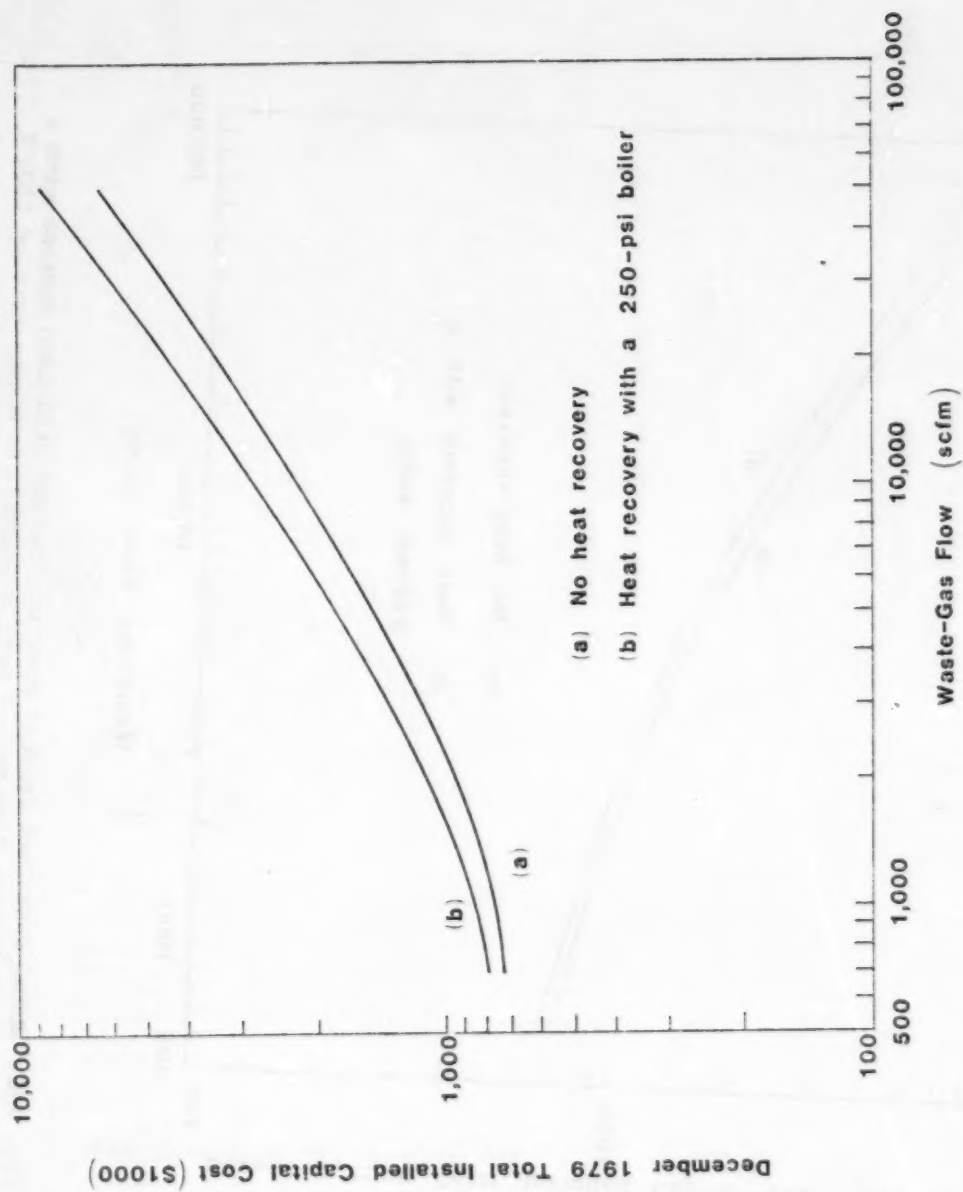


Fig. V-9. Total Installed Capital Cost for Thermal Oxidation Systems with a Scrubber at a Residence Time of 0.5 sec, a Combustion Temperature of 2200 F, and a Waste-Gas Heat Content of 1 Btu/scf

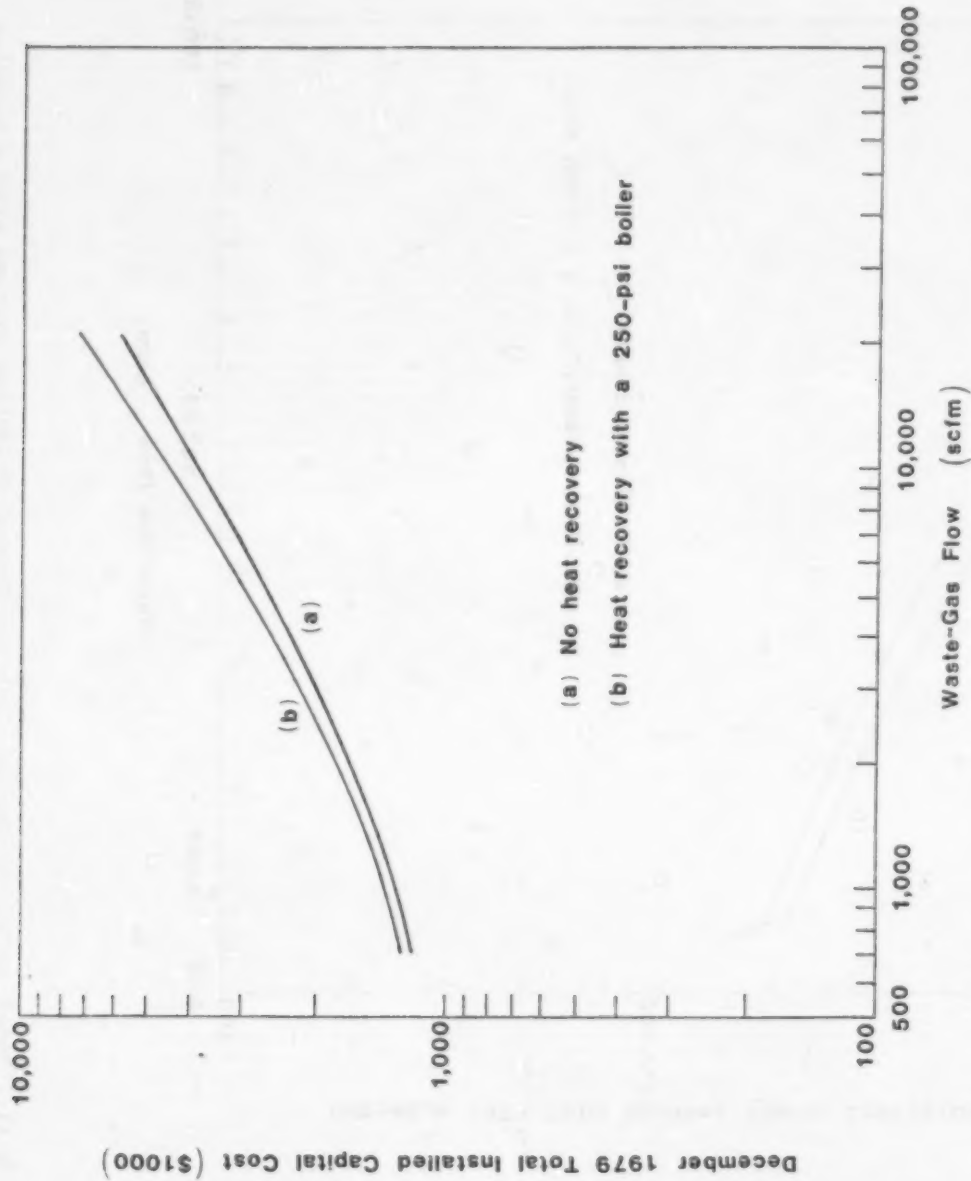


Fig. V-10. Total Installed Capital Cost for Thermal Oxidation Systems with a Scrubber at a Residence Time of 0.5 sec, a Combustion Temperature of 2600 F, and a Waste-Gas Heat Content of 1 Btu/scf

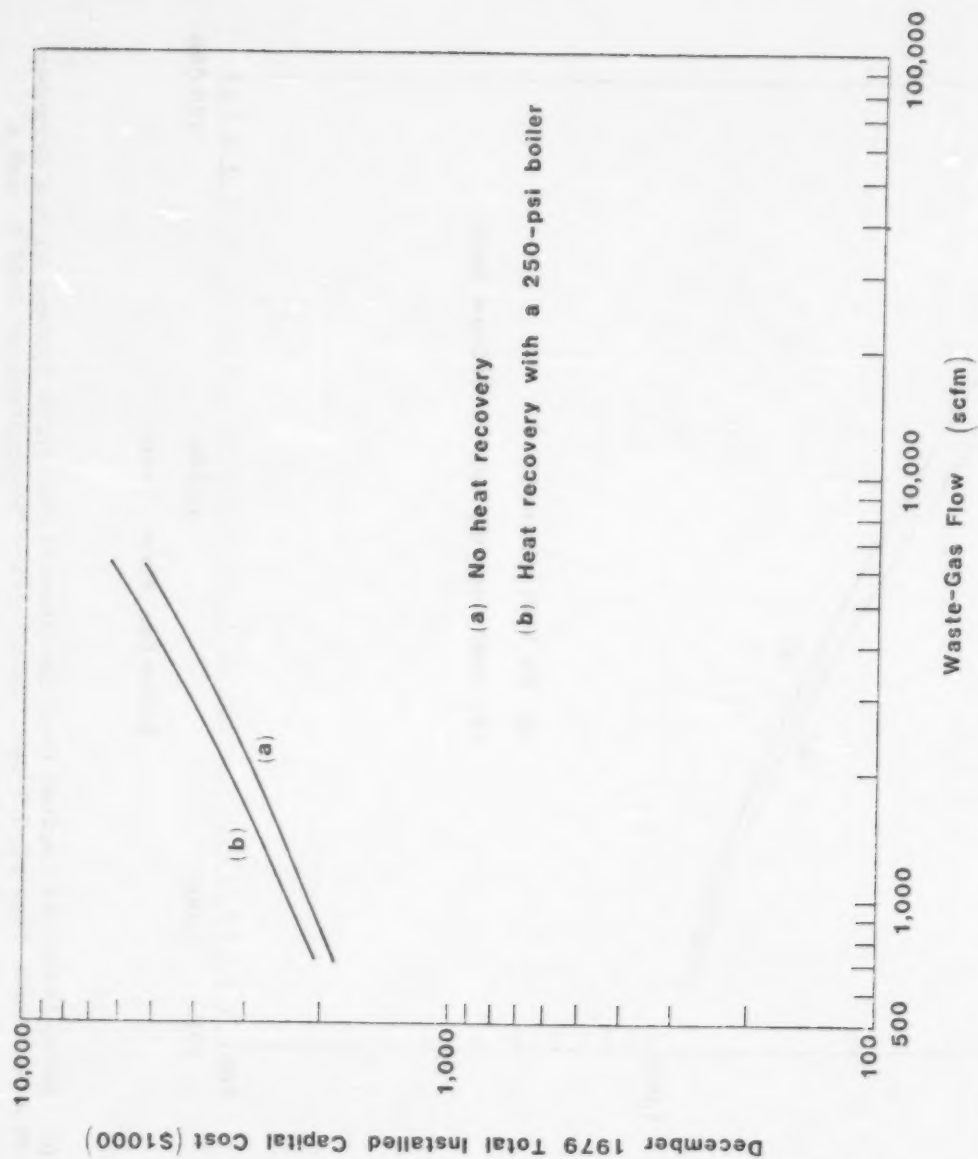


Fig. V-11. Total Installed Capital Cost for Thermal Oxidation Systems with a Scrubber at a Residence Time of 0.5 sec, a Combustion Temperature of 3000°F, and a Waste-Gas Heat Content of 1 Btu/scf

3. Waste-Heat Boiler

Preliminary purchase quotes for steam-generating waste-heat boilers were obtained from vendors for operation at various steam pressure levels and various proportions of heat recovery. The installed costs were then estimated by applying installation factors to these purchase costs as previously described. The installed cost of the 2600°F unit was assumed to be 10% greater than that of the 2200°F unit of the same heat duty, and the cost of the 3000°F was assumed to be 20% greater than that of the 2200°F unit. Multiple units were used when the capacity was in excess of 200,000 lb/hr of steam, or 240,000,000 Btu/hr. The cost curves shown in Figs. V-4 and V-5 for the boilers were developed as described previously.

4. Scrubbers

The estimated installed costs were obtained by applying installation factors to the purchase cost of the column as previously described. The scrubber cost curve shown in Fig. V-1 includes the cost of the quench chamber and the recycle water pump and the additional capital cost involved with a larger fan to overcome the pressure drop across the scrubber.

For ease in evaluating system component combinations, all components except the scrubber are presented as a factor of the total waste-gas flow in scfm. The scrubber flow rate is determined by multiplying the waste-gas flow by the flue-gas ratio (Fig. III-3) and the quench flow ratio (Fig. III-8).

C. ANNUAL COSTS

Annual costs for various operating conditions are presented in Appendix A for sulfur-containing VOC and in Appendix B for halogen-containing VOC. The heat recovery case refers to a 250-psi waste-heat boiler. These costs are the basis for all the cost-effectiveness graphs included in the report. The basis used in calculating these annual costs is defined in Table V-2.

Figures V-12 through V-15 show the annual cost of thermal oxidation for various sulfur-containing VOC cases. Figures V-16 through V-23 show the annual cost of thermal oxidation for various halogen-containing VOC cases. The annual cost increases for higher waste-gas heat contents are due to the use of larger amounts of caustic. Figure V-24 shows the annual cost of thermal oxidation for various combustion temperatures.

Table V-2. Annual Cost Parameters

Operating factor	8760 hr/yr ^a
Operating labor	\$15/man-hr
Fixed costs	
Maintenance labor plus materials, at 6%	} 29% installed capital
Capital recovery, at 18% ^b	
Taxes, insurances, administrative charges, at 5%	
Utilities	
Electric power	\$0.03/kWh
Natural gas	
Heat recovery credit (equals natural gas)	\$2.00/million Btu
Caustic (50% NaOH)	5¢/lb
Makeup water	25¢/1000 gal

^aProcess downtime is normally expected to range from 5 to 15%. If the hourly rate remains constant, the annual production and annual VOC emissions will be correspondingly reduced. Control devices will usually operate on the same cycle as the process. From the standpoint of cost-effectiveness calculations, the error introduced by assuming continuous operation is negligible.

^bBased on 10-year life and 12% interest.

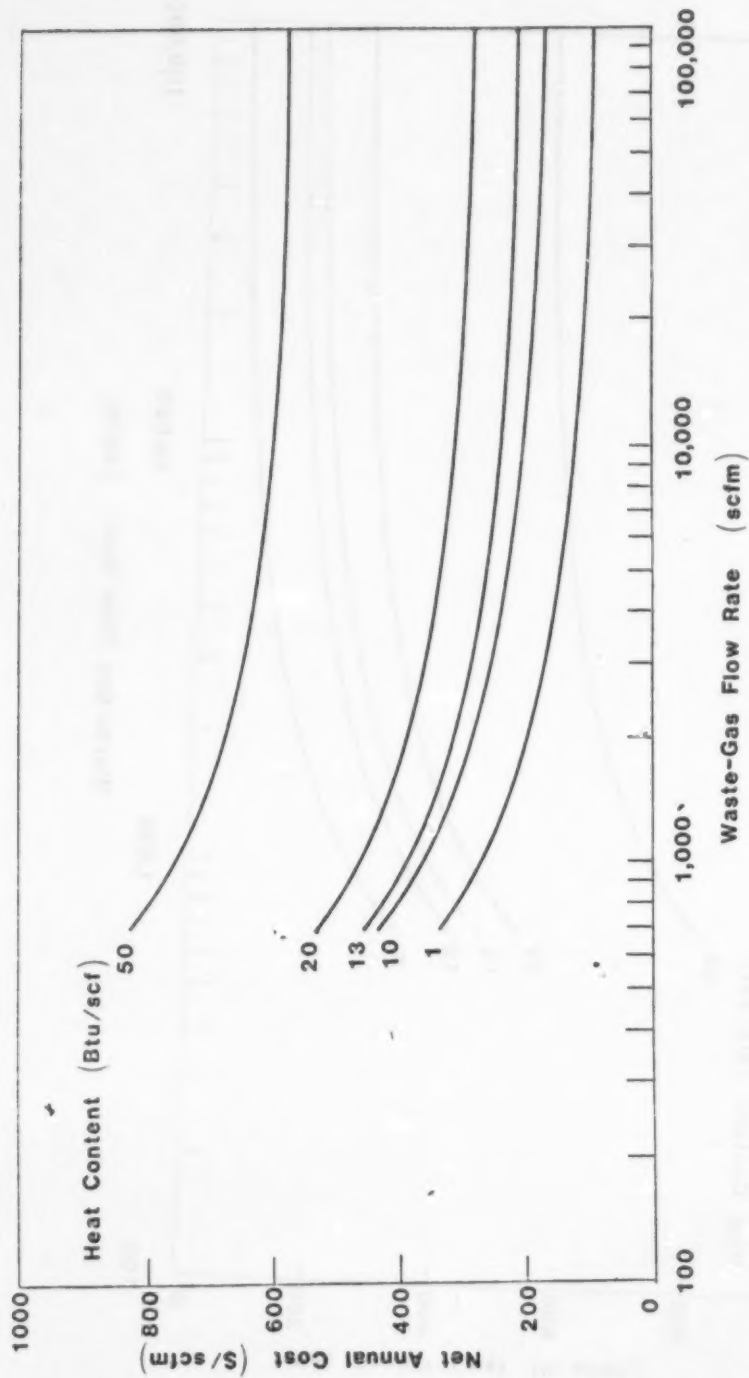


Fig. V-12. Net Annual Costs vs Waste-Gas Flow for Thermal Oxidizers with a Scrubber Using No Heat Recovery, 1400°F Combustion Temperature, 0.5-sec Residence Time, and a Heat Content of from 1 to 50 Btu/scf

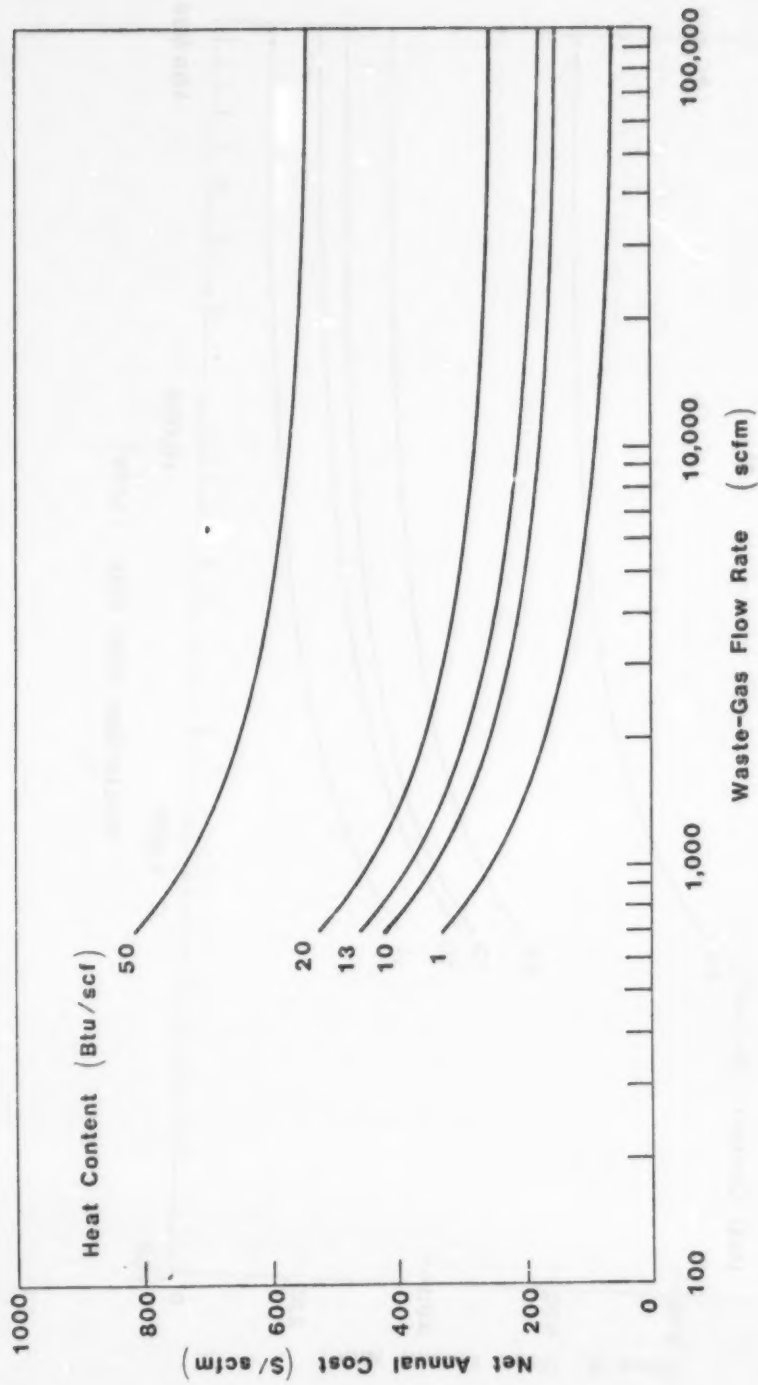


Fig. V-13. Net Annual Costs vs Waste-Gas Flow for Thermal Oxidizers with a Scrubber Using Heat Recovery, 1400°F Combustion Temperature, 0.5-sec Residence Time, and a Heat Content of from 1 to 50 Btu/scf

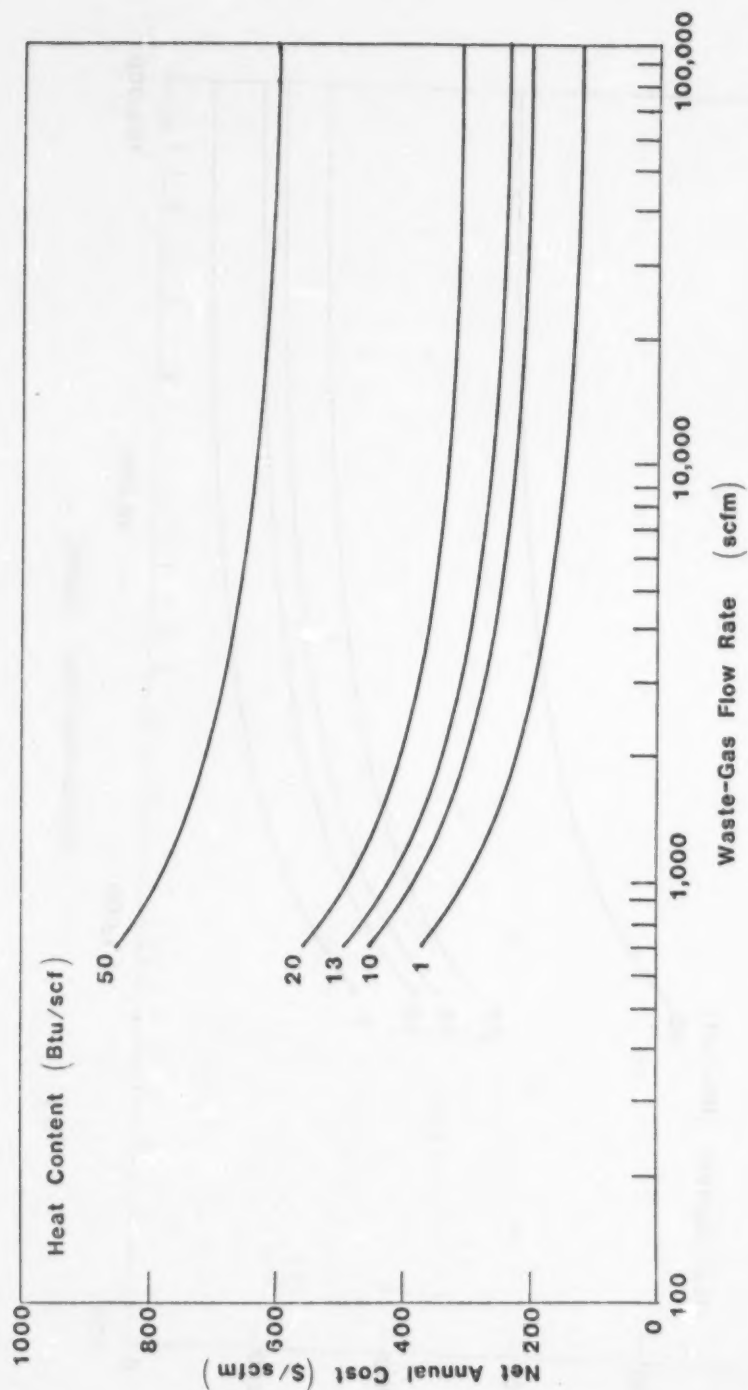


Fig. V-14. Net Annual Costs vs Waste-Gas Flow for Thermal Oxidizers with a Scrubber Using No Heat Recovery, 1600°F Combustion Temperature, 0.5-sec Residence Time, and a Heat Content of from 1 to 50 Btu/scf

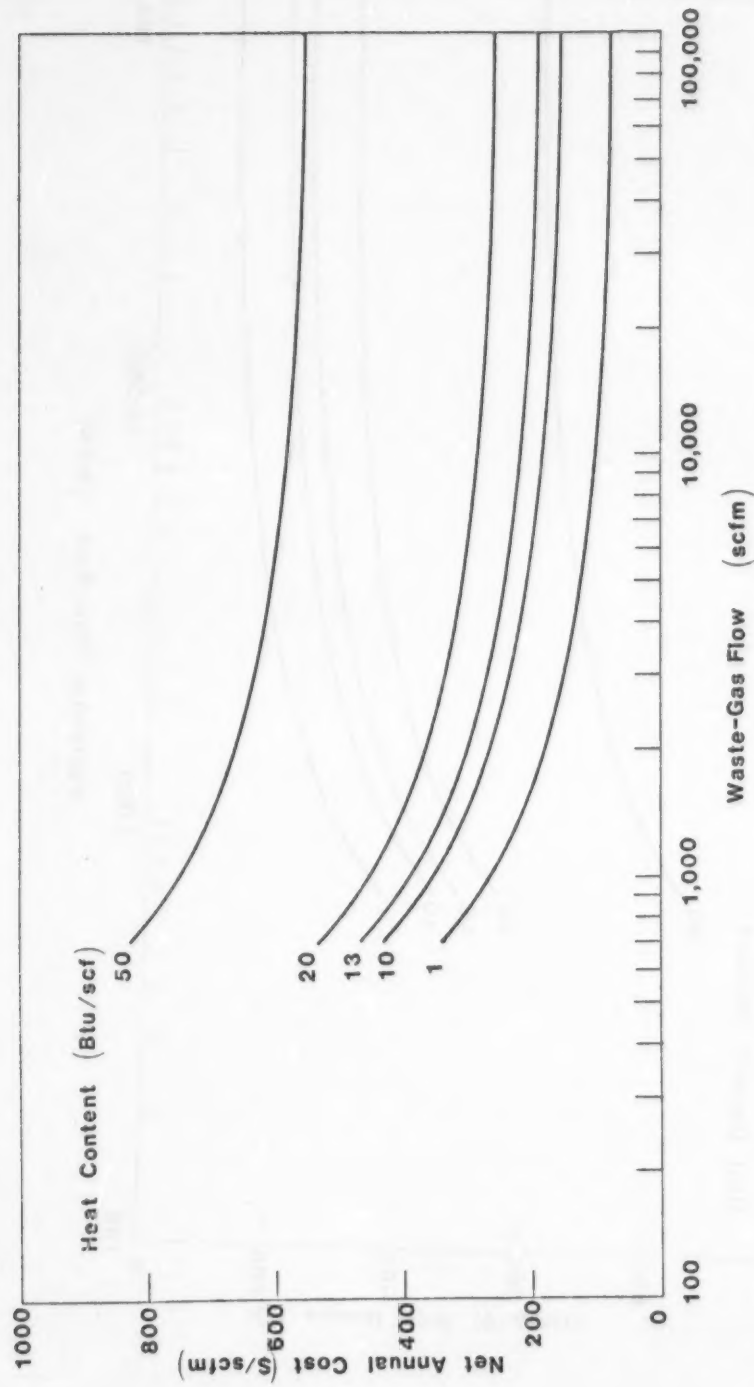


Fig. V-15. Net Annual Costs vs Waste-Gas Flow for Thermal Oxidizers with a Scrubber Using Heat Recovery, 1600°F Combustion Temperature, 0.5-sec Residence Time, and a Heat Content of from 1 to 50 Btu/scf

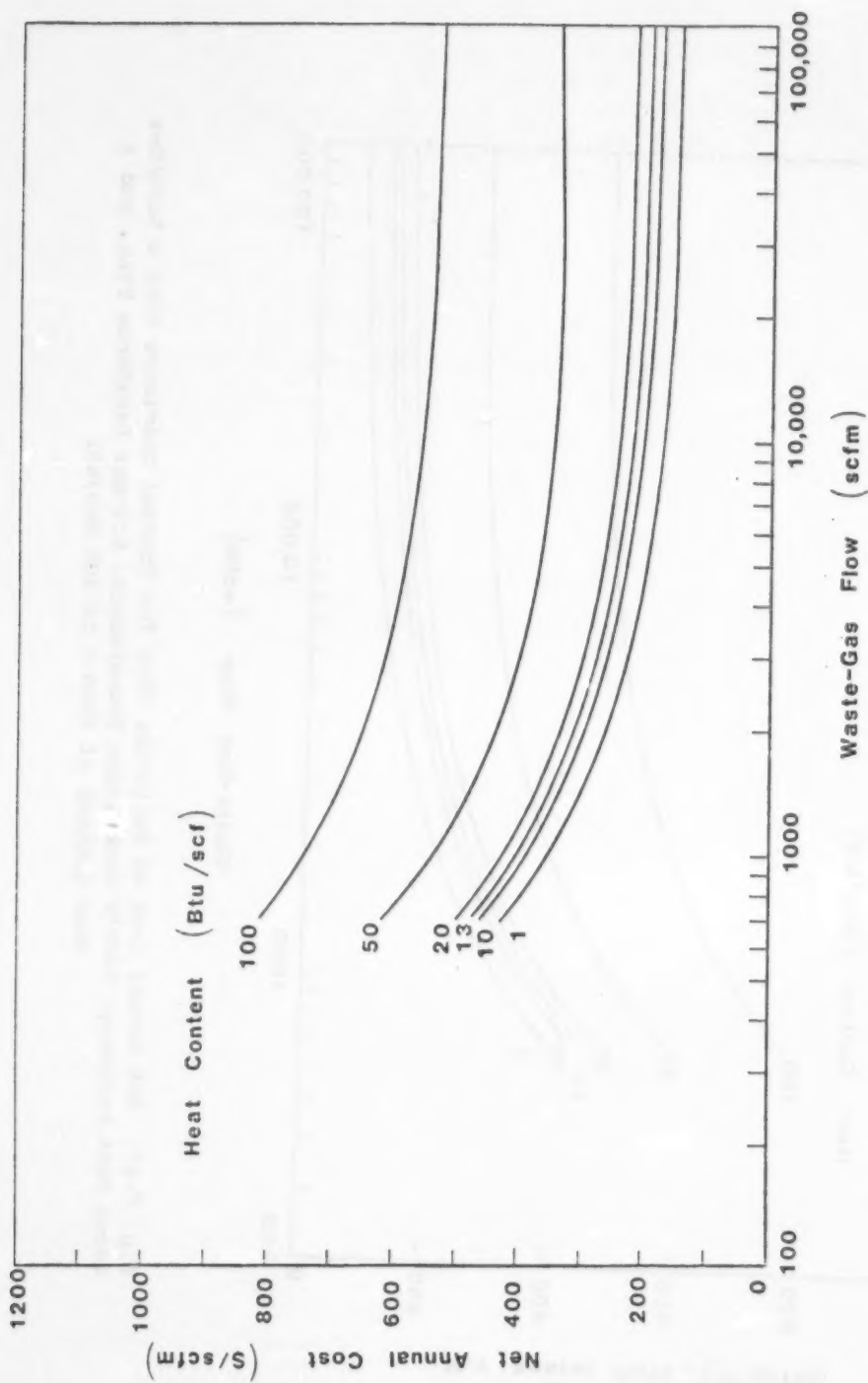


Fig. V-16. Net Annual Cost vs Waste-Gas Flow for Thermal Oxidizers with a Scrubber Using No Heat Recovery, 1800°F Combustion Temperature, 0.5-sec Residence Time, and a Heat Content of from 1 to 100 Btu/scf

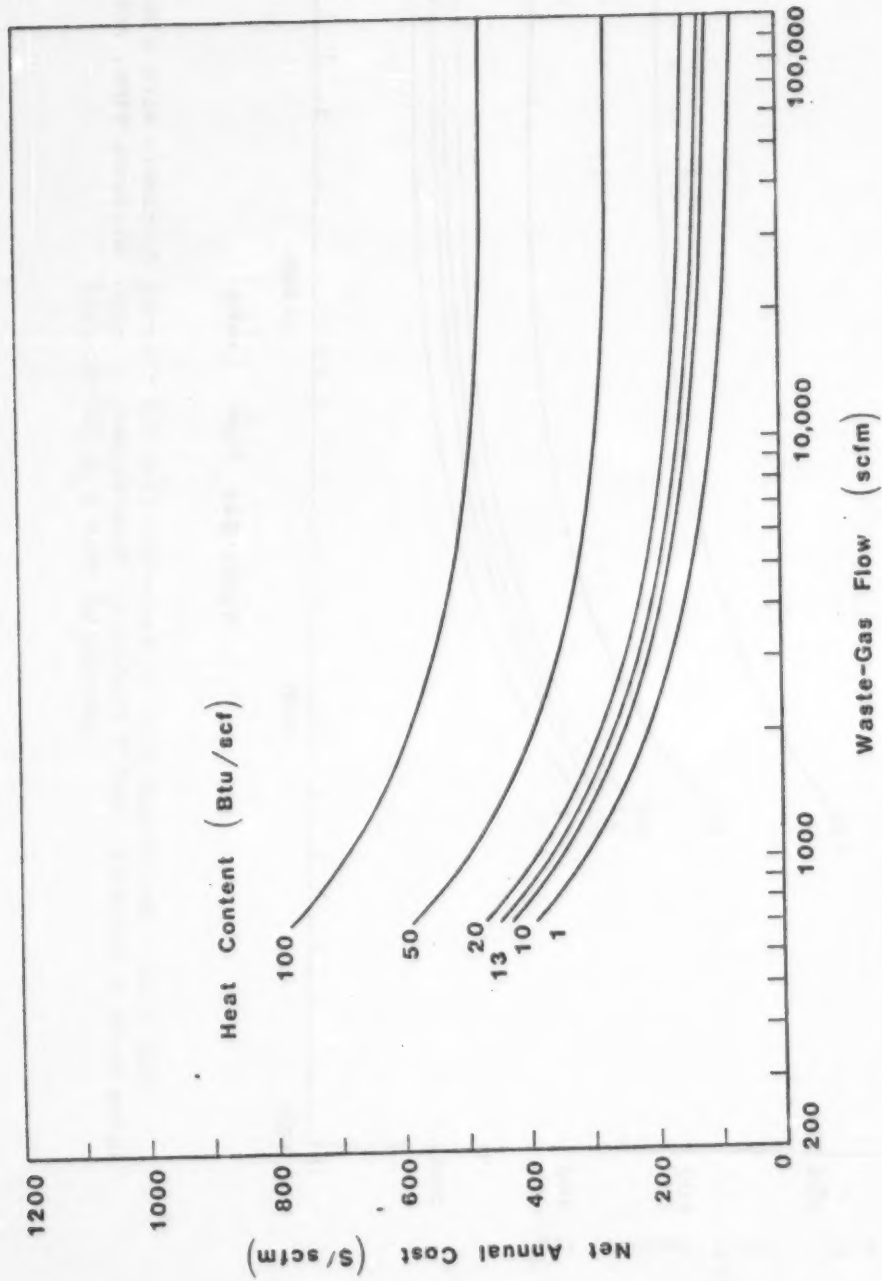


Fig. V-17. Net Annual Cost vs Waste-Gas Flow for Thermal Oxidizers with a Scrubber Using Heat Recovery, 1800°F Combustion Temperature, 0.5-sec Residence Time, and a Heat Content of from 1 to 100 Btu/scf

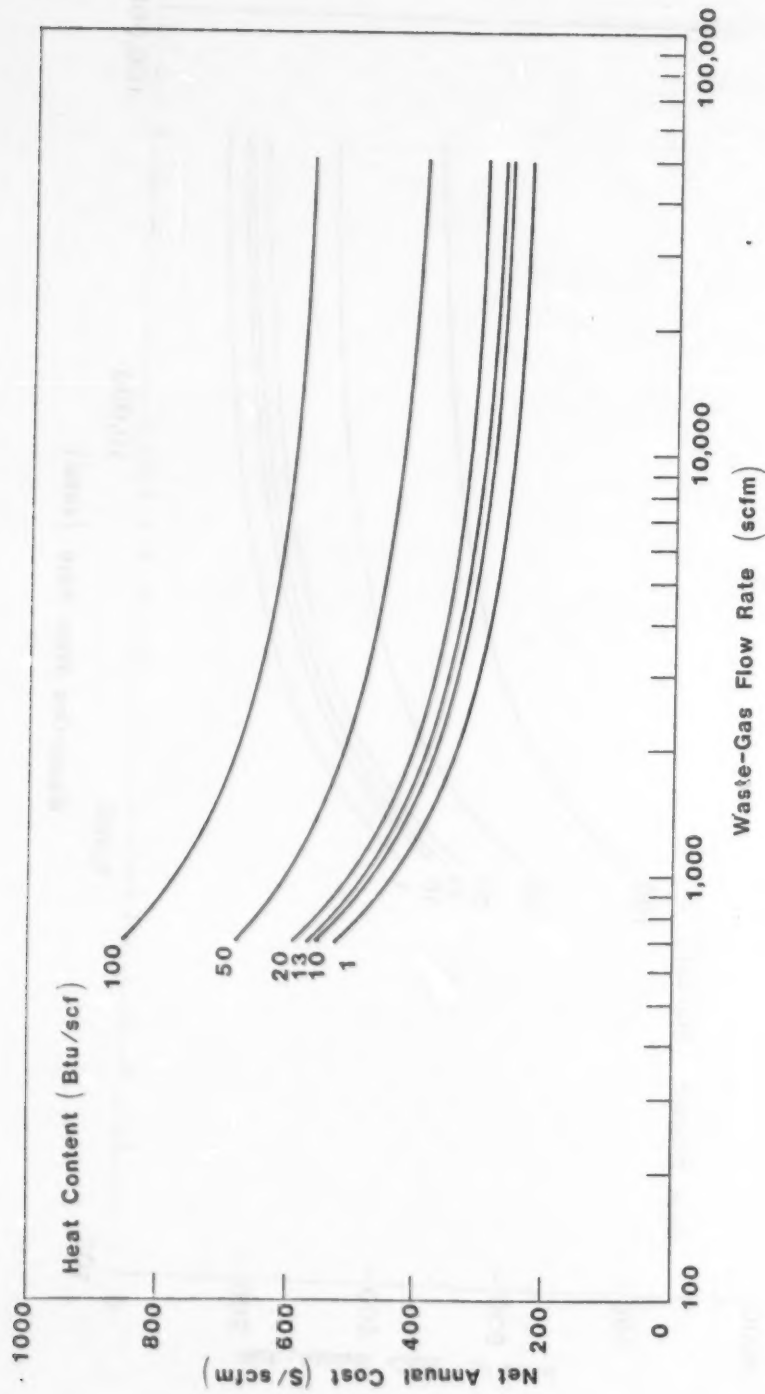


Fig. V-18. Net Annual Costs vs Waste-Gas Flow for Thermal Oxidizers with a Scrubber Using No Heat Recovery, 2200°F Combustion Temperature, 0.5-sec Residence Time, and a Heat Content of from 1 to 100 Btu/scf

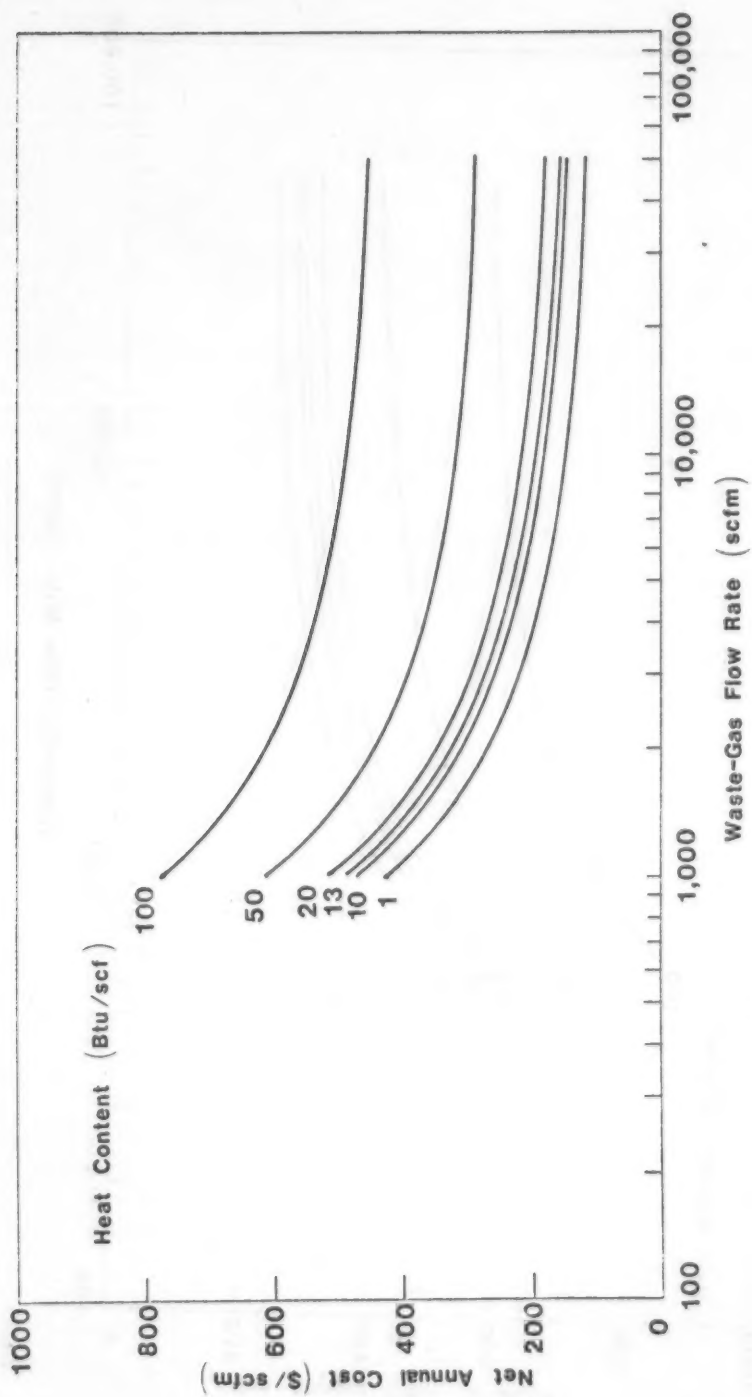


Fig. V-19. Net Annual Costs vs Waste-Gas Flow for Thermal Oxidizers with a Scrubber Using Heat Recovery, 2200°F Combustion Temperature, 0.5-sec Residence Time, and a Heat Content of from 1 to 100 Btu/scf

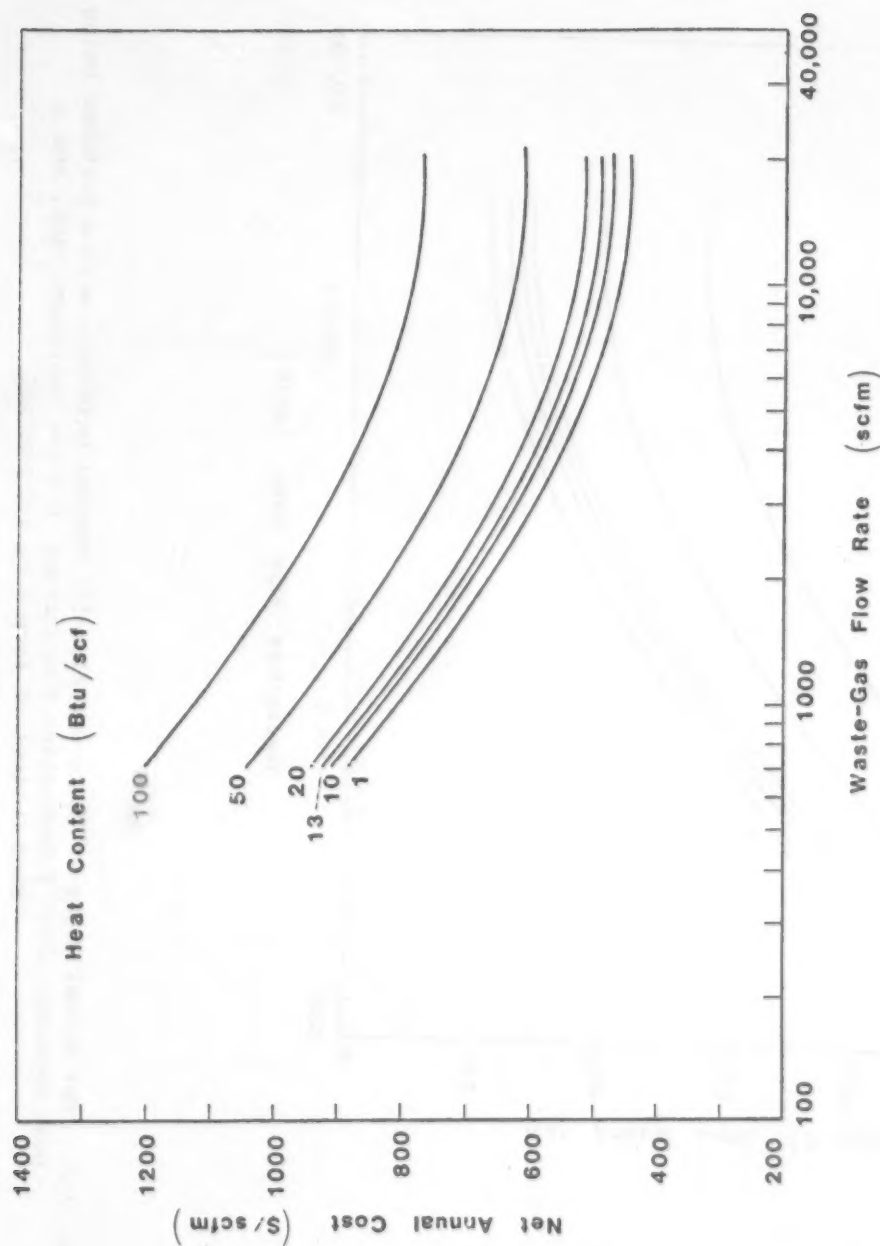


Fig. V-20. Net Annual Costs vs Waste-Gas Flow for Thermal Oxidizers with a Scrubber Using No Heat Recovery, 2600 F Combustion Temperature, 0.5-sec Residence Time, and a Heat Content of from 1 to 100 Btu/scf

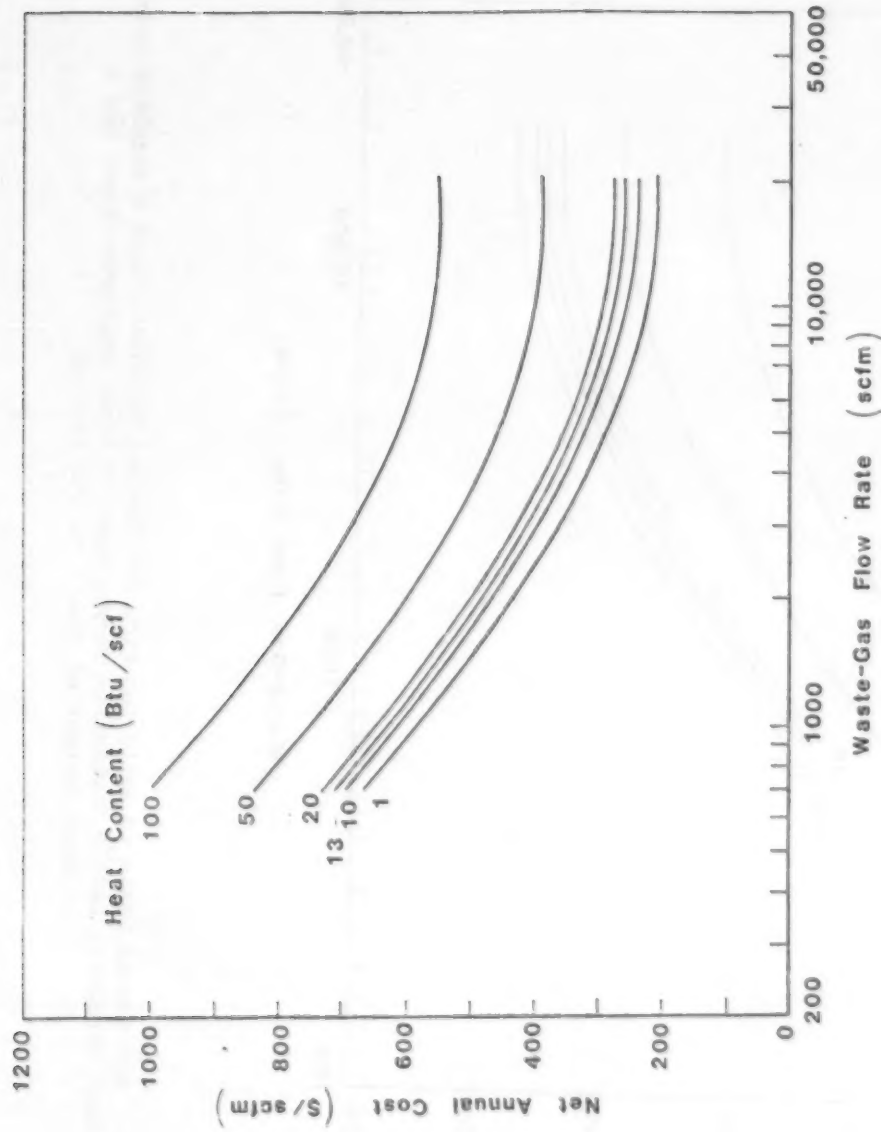
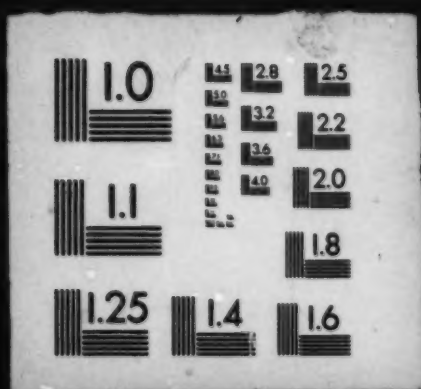


Fig. V-21. Net Annual Costs vs Waste-Gas Flow for Thermal Oxidizers with a Scrubber Using Heat Recovery, 2600°F Combustion Temperature, 0.5-sec Residence Time, and a Heat Content of from 1 to 100 Btu/scf

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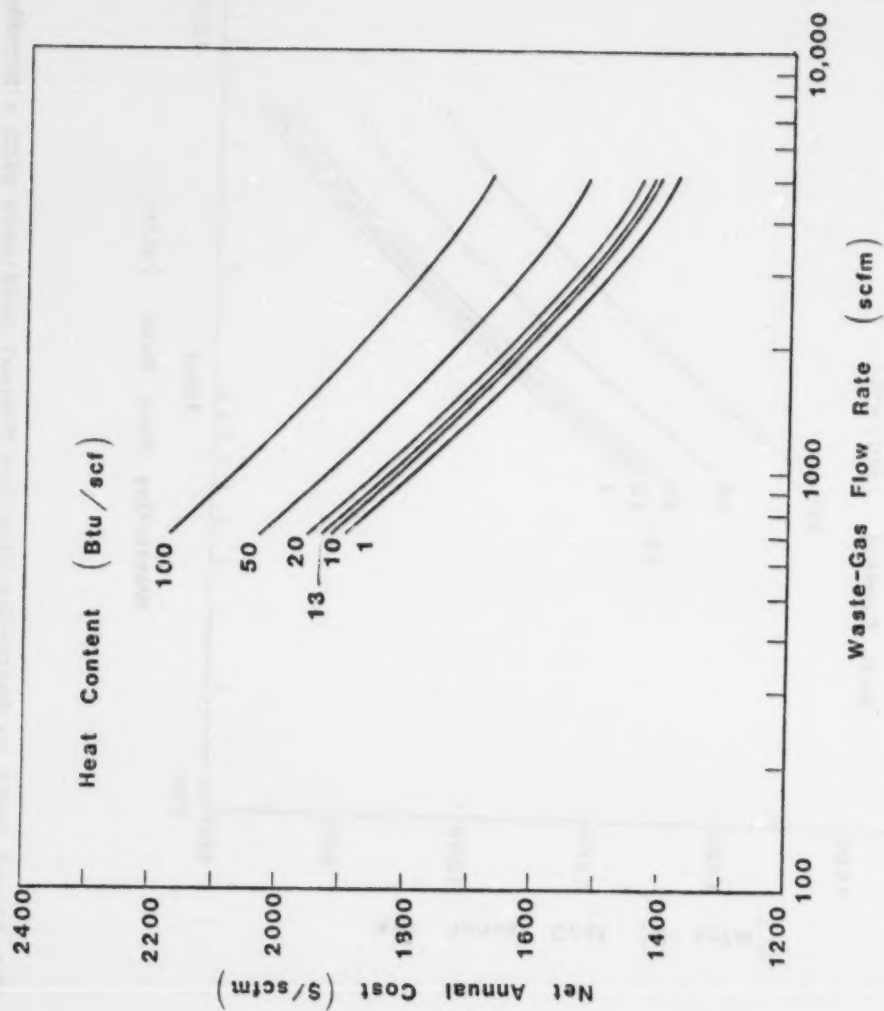


Fig. V-22. Net Annual Costs vs Waste-Gas Flow for Thermal Oxidizer with a Scrubber Using No Heat Recovery, 3000°F Combustion Temperature, 0.5-sec Residence Time, and a Heat Content of from 1 to 50 Btu/scf

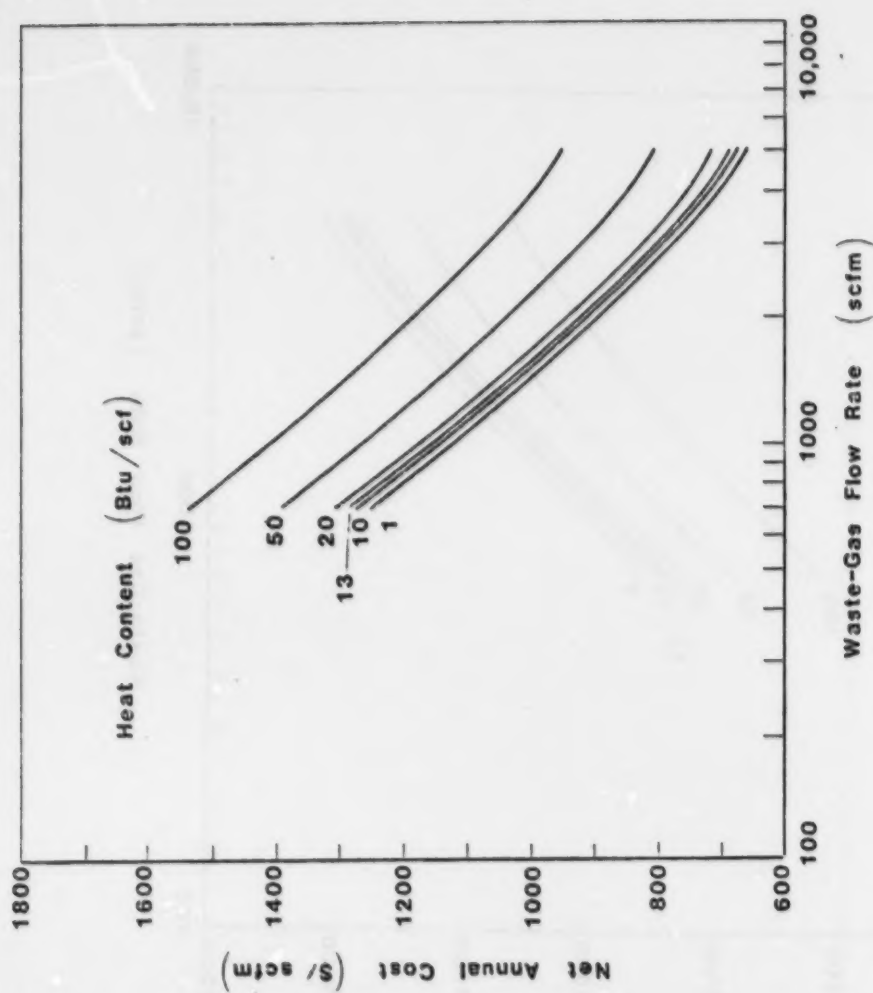


Fig. V-23. Net Annual Costs vs Waste-Gas Flow for Thermal Oxidizers with a Scrubber Using Heat Recovery, 3000°F Combustion Temperature, 0.5-sec Residence Time, and a Heat Content of from 1 to 100 Btu/scf

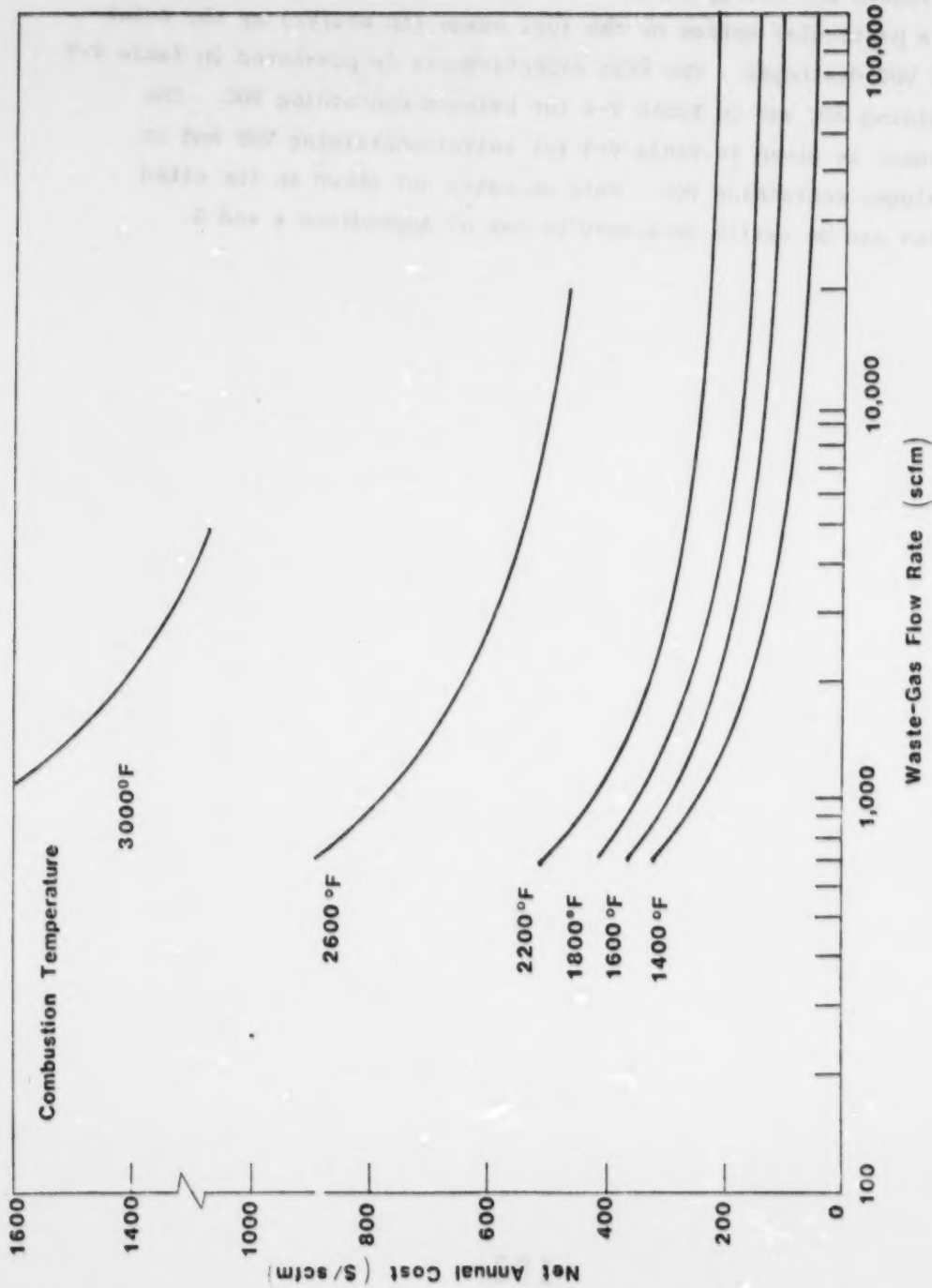


Fig. V-24. Net Annual Cost vs Waste-Gas Flow for Thermal Oxidizers with a Scrubber Using No Heat Recovery, 0.5-sec Residence Time, Heat Content of 1 Btu/scf, and a Combustion Temperature of from 1400°F to 3000°F

D. COST EFFECTIVENESS AND ENERGY EFFECTIVENESS

The cost effectiveness and energy effectiveness are calculated by dividing the annual cost for a particular option or the fuel usage (in Btu/yr) by the total annual amount of VOC destroyed. The cost effectiveness is presented in Table V-3 for sulfur-containing VOC and in Table V-4 for halogen-containing VOC. The energy effectiveness is given in Table V-5 for sulfur-containing VOC and in Table V-6 for halogen-containing VOC. Data on cases not shown in the cited tables and figures can be easily developed by use of Appendices A and B.

Table V-3. Cost Effectiveness of Thermal Oxidation and Scrubbing for Control of Sulfur-Containing VOC

Waste-Gas Heat Content (Btu/scf)	Waste- Gas Flow Rate (scfm)	VOC Destroyed (lb/hr) ^c at 90%	VOC Destroyed (lb/hr) ^c at 99%	Cost Effectiveness (per lb of VOC Destroyed)			
				90% VOC Destruction ^a		99% VOC Destruction ^b	
				No Heat Recovery	Waste Heat Boiler 250- psi Steam	No Heat Recovery	Waste Heat Boiler 250- psi Steam
1	700	5.06	5.57	\$5.35	\$5.26	\$5.27	\$4.90
	5,000	36.2	39.8	2.24	1.73	2.39	1.73
	20,000	144.7	159.2	1.73	1.20	1.99	1.24
	50,000	361.8	398.0	1.58	1.00	1.76	1.03
	100,000	723.4	796.0	1.50	0.92	1.68	0.95
10	700	50.6	55.7	0.67	0.66	0.65	0.61
	5,000	362	398	0.36	0.31	0.37	0.30
	20,000	1,447	1,592	0.31	0.26	0.33	0.25
	50,000	3,618	3,980	0.30	0.24	0.30	0.23
	100,000	7,234	7,960	0.29	0.23	0.30	0.22
13	700	65.8	72.4	0.56	0.55	0.54	0.51
	5,000	471	517	0.32	0.28	0.31	0.26
	20,000	1,881	2,070	0.28	0.24	0.28	0.23
	50,000	4,703	5,174	0.26	0.22	0.27	0.21
	100,000	9,404	10,348	0.26	0.21	0.26	0.20
20	700	101.2	111.4	0.42	0.41	0.40	0.38
	5,000	724	796	0.26	0.23	0.25	0.22
	20,000	2,894	3,184	0.23	0.21	0.23	0.20
	50,000	7,236	7,960	0.23	0.20	0.22	0.19
	100,000	14,472	15,920	0.22	0.19	0.22	0.18
50	700	253	278.5	0.26	0.26	0.24	0.24
	5,000	1,810	1,990	0.20	0.19	0.19	0.17
	20,000	7,235	7,960	0.19	0.18	0.18	0.16
	50,000	18,060	19,900	0.18	0.17	0.17	0.16
	100,000	36,170	39,800	0.18	0.17	0.17	0.16

^aBased on 1400°F combustion temperature and 0.5-sec residence time.^bBased on 1600°F combustion temperature and 0.5-sec residence time.^cVOC molecular weight = 100; molar heat of combustion = 746,000 Btu/lb-mole of VOC.

Table V-4. Cost Effectiveness of Thermal Oxidation and Scrubbing for Control of Halogen-Containing VOC

Cost Effectiveness (per lb of VOC Destroyed)								
Waste-Gas Heat Content (Btu/scf)	Waste-Gas Flow Rate (scfm)	VOC Destroyed (lb/hr) ^a at 99.9	Combustion Temperature 1800°F		Combustion Temperature 2200°F		Combustion Temperature 2600°F	
			No Heat Recovery	Waste Heat Boiler 250-psi Steam	No Heat Recovery	Waste Heat Boiler 250-psi Steam	No Heat Recovery	Waste Heat Boiler 250-psi Steam
1	700	5.62	\$6.03	\$5.53	\$7.41	\$6.30	\$12.50	9.46
	5,000	40.2	2.73	1.81	3.95	2.54	7.35	4.11
	20,000	160.8	2.22	1.22	3.43	1.97	6.33	3.13
	50,000	402	2.06	1.07	3.24	1.66		
10	700	56.2	0.65	0.60	0.78	0.67	1.29	0.99
	5,000	402	0.32	0.23	0.44	0.30	0.78	0.45
	20,000	1,608	0.27	0.17	0.39	0.24	0.68	0.36
	50,000	4,020	0.26	0.16	0.37	0.21		
13	700	73.1	0.51	0.48	0.61	0.53	1.00	0.77
	5,000	523	0.26	0.19	0.35	0.24	0.61	0.36
	20,000	2,090	0.22	0.15	0.31	0.20	0.53	0.28
	50,000	5,230	0.21	0.13	0.29	0.17		
20	700	112.4	0.35	0.33	0.42	0.36	0.67	0.52
	5,000	804	0.19	0.14	0.24	0.17	0.41	0.25
	20,000	3,216	0.16	0.11	0.22	0.14	0.36	0.20
	50,000	8,040	0.16	0.11	0.21	0.13		
50	700	281	0.17	0.17	0.20	0.17	0.30	0.23
	5,000	2,010	0.11	0.10	0.13	0.10	0.19	0.19
	20,000	8,040	0.10	0.08	0.12	0.09	0.17	0.17
	50,000	20,100	0.10	0.08	0.11	0.08		
100	700	562	0.12	0.11	0.12	0.11	0.17	0.14
	500	4,020	0.08	0.07	0.09	0.07	0.12	0.09
	20,000	16,080	0.08	0.07	0.08	0.07	0.11	0.08
	50,000	40,200	0.08	0.07	0.08	0.06		

^aVOC molecular weight = 100; molar heat of combustion = 746,000 Btu/lb-mole of VOC.

Table V-5. Fuel Energy Effectiveness of Thermal Oxidation
for Control of Sulfur-Containing VOC

Waste-Gas Heat Content (Btu/scf)	Fuel Energy Usage (Btu/scf)		Net Energy Usage ^a (Btu/scf)		Energy Effectiveness ^d (Btu/lb of VOC Destroyed)			
	90% VOC ^b Destruction	99% VOC ^c Destruction	90% VOC ^b Destruction	99% VOC ^c Destruction	90% VOC Destruction ^b No Heat Recovery	Waste Heat- Boiler 250- psi Steam	99% VOC Destruction ^c No Heat Recovery	Waste Heat Boiler 250- psi Steam
1	63.5	82.3	26.3	30.6	527,100	218,300	620,600	230,700
10	54.2	73.0	17.0	21.3	45,000	14,100	55,000	16,100
13	51.0	69.9	13.8	18.2	32,600	8,800	40,500	10,600
20	43.6	62.5	6.37	10.8	18,100	2,600	23,600	4,100
50	12.0	31.1	(25.2)	(20.6)	2,000	(4,200)	4,700	(3,100)

^aBased on a waste heat boiler generating 250-psi steam.

^bBased on 1400°F combustion temperature and 0.5-sec residence time.

^cBased on 1600°F combustion temperature and 0.5-sec residence time.

^dVOC molecular weight = 100; molar heat of combustion = 746,400 Btu/lb-mole of VOC.

Table V-6. Fuel Energy Effectiveness of Thermal Oxidation
for Control of Halogen-Containing VOC

Waste-Gas Heat Content (Btu/scf)	Energy Effectiveness (Btu/lb of VOC Destroyed) ^a					
	Combustion Temperature 1800°F		Combustion Temperature 2200°F		Combustion Temperature 2600°F	
	No Heat Recovery	Waste Heat Boiler 250- psi Steam	Heat Recovery	Waste Heat Boiler 250- psi Steam	No Heat Recovery	Waste Heat Boiler 250- psi Steam
1	777,200	247,900	1,225,600	306,400	2,466,200	739,900
10	71,000	18,100	115,800	23,900	246,600	62,800
13	52,800	12,200	85,700	14,900	175,900	43,100
20	31,800	5,300	53,100	7,100	112,100	25,800
50	8,200	(2,400)	14,000	(4,300)	36,800	2,200
100	400	(4,900)	1,700	(7,500)	12,600	(4,700)

^aVOC molecular weight = 100; molar heat of combustion = 746,400 Btu/lb-mole of VOC.

VI. SUMMARY AND CONCLUSIONS

Thermal oxidation is a widely used control technique for control of sulfur-containing and halogen-containing VOC. This evaluation describes the limits and design principles of this technique. Design criterion and design procedures are presented that allow for the preliminary design of high-temperature thermal oxidizers and flue-gas scrubbing for both conventional and high-temperature thermal oxidizers. Thermal oxidizers without heat recovery and with waste-heat steam-boiler heat recovery are considered. Capital and operating costs are developed and the annual cost of thermal oxidation is calculated as a function of the characteristics of the waste gas.

The conclusions derived from the cost evaluation are as follows:

1. Since the thermal oxidizer design used here is quite conservative, the control costs actually experienced in industry are expected to be less than the costs presented in this report.
2. The waste-gas heat content (VOC content) is a highly sensitive variable in determining annual costs, cost effectiveness, and energy effectiveness. As the heat content of the waste gas increases, the annual cost and the cost per scf increase, whereas the energy effectiveness and the cost effectiveness sharply improve. The increase in annual cost and cost per cfm is due to the increased amounts of caustic and water required to control the SO_2 or HCl in the flue gas. This is largely the result of the assumption that the VOC in the waste gas is 100% sulfonated or halogenated. When the sulfur-containing or halogen-containing VOC is actually a small percentage of the VOC in the waste gas, the caustic and water requirements will be less than those presented in this report.
3. The waste-gas flow rate is a highly sensitive variable in determining annual costs and cost effectiveness; energy effectiveness is independent of the flow rate. As the waste gas flow increases (at a constant heat content), the annual costs increase but the annual cost per scfm of waste gas and the cost per lb of VOC decrease. This ratio decreases drastically between low flows (700 scfm)

and moderate flows (5000 scfm), but remains relatively constant between moderate (5000 scfm) to large flows (50,000 to 100,000 scfm). Energy effectiveness per scf is constant with flow.

4. Net annual costs for controlling the VOC decrease when heat recovery is included, provided that there is a use for the steam generated.

5. For high-temperature thermal oxidation (1800 to 3000°F) the annual cost, cost effectiveness, and energy effectiveness increase exponentially with an increase in combustion temperature.

6. The effects of residence time, VOC destruction efficiency for conventional thermal oxidation, and different heat recovery systems on annual costs, cost effectiveness, and energy effectiveness can be found in the control-device evaluation report for thermal oxidation¹.

VII. REFERENCES

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APPENDIX A
ANNUAL COST DATA FOR SULFUR-CONTAINING VOC CONTROL

Annual Cost Data Calculations
(Applies to Appendices A and B)

The annual costs of thermal oxidation systems are presented in Tables A-5 to A-14 and B-3 to B-20. Each table shows the costs at a specific offgas heat content and combustion temperature at various flow rates, residence time, and with and without heat recovery. The heat recovery case includes a 250-psi waste heat boiler.

The following sample calculation is for a stream with an offgas heat content of 10 Btu/scf, a combustion temperature of 2600°F, a residence time of 0.5 sec, a flow rate of 5000 scfm, and with heat recovery.

Capital cost = thermal oxidizer (\$1,250,000) + waste heat boiler (\$734,000) + scrubber (\$912,000) = \$2,896,000

Thermal oxidizer = \$1,250,000 -- from Fig. V-3 at 5000 scfm, 0.5 sec resistance, and 2600°F

Waste heat boiler = \$734,000 -- from Fig. V-5 at 5000 scfm, 250 psi, and 2600°F

Scrubber = \$912,000 -- from Fig. V-1 at 5000 scfm X 5.1 (from Fig. III-3) X 1.07 (from Fig. III-8)

Fixed cost = \$2,896,000 X 0.29* = \$840,000/yr

Operating cost = fuel (\$1,660,000) + electricity (\$23,100) + makeup water quench (\$1,300) + makeup water scrubber (\$10,200) + caustic (\$243,200) + labor (\$36,000) = \$1,974,000/yr

Fuel = (315.8 Btu/scf (from figure III-2)) X (5000 scfm) X (\$2.00/million Btu*) X (60 min/hr) X (8,760 hr/yr) = \$1,660,000/yr

Electricity = (22 in H₂O) X (.000157 $\frac{\text{hp}}{\text{in. H}_2\text{O}}$) X (1/.60 efficiency) X (.746 kWh) X (5000 scfm) X (\$0.03/kWh*) X (8760 hr/yr) = \$5,640/yr plus electricity for combustion air blower = 3.1 (from table III-3) X \$5,640/yr = \$17,500/yr + \$5,640/yr = \$23,100/yr

* See table V-2.

Makeup water quench = (2 gal/1000 scf (from Fig. III-9)) X (5000 scfm) X

(\$0.25/1000 gal*) X (60 min/hr) X (8,760 hr/yr) = \$1,300/yr

Makeup water scrubber = [19.2 gal/lb of chlorine (from text III-B-5)] X

(0.0487 lb/hr of chlorine/scfm) X (5000 scfm) X (\$0.25/1000 gal*) X

(8,760 hr/yr) = \$10,200/yr

Caustic = [1.14 lb/lb of chlorine (from text III-B-5)] X (0.0487 lb/hr of

chlorine/scfm) X (5000 scfm) X (\$0.10/lb of 100% caustic*) X (8,760 hr/yr) =

\$243,200/yr

Labor = \$36,000/yr (from ref 1); the labor cost for a system without heat

recovery is \$18,000/yr.

Credit = (13,900 Btu/hr/scfm) X (5000 scfm) X (\$2.0/million Btu*) X (8,760 hr/yr) =

\$1,218,000/yr

Annual cost = fixed cost (\$840,000) + operating cost (\$1,974,000) - credit (\$1,218,000)

\$1,596,000

Net cost = $\frac{\text{Annual cost } (\$1,596,000/\text{yr})}{\text{Flow rate (5000 scfm)}} = 319.22 \text{ \$/scfm}$

OFF-GAS HEAT CONTENT 1.00 BTU/SCF
COMBUSTION TEMPERATURE 1400 F

CASE	OFF-GAS FLOW SCFH	CAPITAL COST (000)	FIXED COST (000)	OPERATING COST (000)	RECOVERY CREDIT (000)	NET ANNUALIZED COST OR CREDIT (000)	NET COST OR SAVINGS PER SCFH \$/SCFH
RESIDENCE TIME 0.50SEC NO HEAT RECOVERY	700. 5000. 20000. 50000. 100000.	567. 1039. 2080. 3703. 5755.	154. 301. 603. 1074. 1659.	73. 413. 1598. 3948. 7918.	0. 0. 0. 0. 0.	238. 714. 2201. 5042. 9587.	339.61 142.85 116.66 100.85 95.87
RESIDENCE TIME 0.50SEC HEAT RECOVERY	700. 5000. 20000. 50000. 100000.	585. 1106. 2471. 4127. 6618.	170. 321. 717. 1197. 1919.	91. 428. 1605. 3959. 7883.	28. 198. 792. 1980. 3960.	233. 551. 1530. 4176. 5843.	332.66 110.22 76.50 63.53 58.43
RESIDENCE TIME 0.75SEC NO HEAT RECOVERY	700. 5000. 20000. 50000. 100000.	568. 1074. 2258. 4022. 6321.	165. 311. 655. 1166. 1833.	73. 413. 1598. 3948. 7918.	0. 0. 0. 0. 0.	238. 724. 2253. 5134. 9751.	340.03 141.89 112.63 102.69 97.51
RESIDENCE TIME 0.75SEC HEAT RECOVERY	700. 5000. 20000. 50000. 100000.	586. 1111. 2449. 4114. 6718.	170. 331. 768. 1289. 2083.	91. 428. 1605. 3959. 7883.	28. 198. 792. 1980. 3960.	233. 561. 1582. 4268. 6007.	333.07 112.25 79.08 63.37 60.07

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OFFGAS HEAT CONTENT 10.00 BTU/SCF
COMBUSTION TEMPERATURE 1400 F

CASE	OFFGAS FLOW SCFM	CAPITAL COST (000)	FIXED COST (000)	OPERATING COST-OR-CREDIT OPERATING COST (000)	RECOVERY CREDIT (000)	NET ANNUALIZED COST OR CREDIT (000)	NET COST OR SAVINGS PLK SCFM \$/SCFM
RESIDENCE TIME 0.50SEC NO HEAT RECOVERY	700. 5000. 20000. 50000. 100000.	567. 1039. 2080. 3705. 5755.	164. 301. 603. 1074. 1669.	137. 868. 3419. 8520. 17622.	0. 0. 0. 0. 0.	301. 1169. 4022. 9594. 18591.	430.55 233.90 201.10 191.89 186.51
RESIDENCE TIME 0.50SEC HEAT RECOVERY	700. 5000. 20000. 50000. 100000.	585. 1106. 2471. 4127. 6618.	170. 321. 717. 1197. 1919.	155. 884. 3426. 8511. 16987.	28. 198. 792. 1980. 3960.	297. 1006. 3351. 7728. 14946.	423.70 201.26 167.54 151.57 149.46
RESIDENCE TIME 0.75SEC NO HEAT RECOVERY	700. 5000. 20000. 50000. 100000.	568. 1074. 2258. 4022. 6321.	165. 311. 655. 1166. 1833.	137. 868. 3419. 8520. 17622.	0. 0. 0. 0. 0.	302. 1180. 4074. 9686. 18855.	431.07 235.93 203.68 193.72 188.55
RESIDENCE TIME 0.75SEC HEAT RECOVERY	700. 5000. 20000. 50000. 100000.	586. 1141. 2649. 4444. 7184.	170. 331. 768. 1289. 2083.	155. 884. 3426. 8511. 16987.	28. 198. 792. 1980. 3960.	297. 1016. 3402. 7820. 15110.	424.11 204.29 170.12 154.41 151.10

OFFGAS HEAT CONTENT 13.00 BTU/SCF
COMBUSTION TEMPERATURE 1400 F

CASE	OFFGAS FLOW SCFH	CAPITAL COST (000)	FIXED COST (000)	OPERATING COST-OR-CREDIT COST (000)	RECOVERY CREDIT (000)	NET ANNUALIZED COST OR CREDIT (000)	NET COST OR SAVINGS PFR SCFH \$/SCFH
RESIDENCE TIME 0.50SEC NO HEAT RECOVERY	700. 5000. 20000. 50000. 100000.	567. 1039. 2080. 3705. 5755.	164. 301. 603. 1074. 1689.	158. 1020. 4026. 10037. 20056.	0. 0. 0. 0. 0.	323. 1321. 4629. 11112. 21725.	461.60 269.24 231.44 222.23 217.25
RESIDENCE TIME 0.50SEC HEAT RECOVERY	700. 5000. 20000. 50000. 100000.	585. 1106. 2471. 4127. 6618.	170. 321. 717. 1197. 1919.	176. 1035. 4033. 10029. 20021.	28. 198. 792. 1980. 3960.	318. 1158. 3958. 9216. 17981.	454.04 231.60 197.88 189.91 179.81
RESIDENCE TIME 0.75SEC NO HEAT RECOVERY	700. 5000. 20000. 50000. 100000.	568. 1074. 2258. 4022. 6321.	165. 311. 655. 1166. 1833.	158. 1020. 4026. 10037. 20056.	0. 0. 0. 0. 0.	323. 1331. 4680. 11204. 21889.	461.41 266.27 234.02 224.07 218.89
RESIDENCE TIME 0.75SEC HEAT RECOVERY	700. 5000. 20000. 50000. 100000.	586. 1141. 2649. 4144. 7184.	170. 331. 768. 1289. 2083.	176. 1035. 4033. 10029. 20021.	28. 198. 792. 1980. 3960.	318. 1168. 4069. 9338. 18145.	454.46 231.63 206.47 184.75 181.45

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OFFGAS HEAT CONTENT 20.00 BTU/SCF
COMBUSTION TEMPERATURE 1400 F

CASE	OFFGAS FLOW SCFM	CAPITAL COST (000)	FIXED COST (000)	OPERATING COST-OR-CREDIT OPERATING COST (000)	RECOVERY CREDIT (000)	NET ANNUALIZED COST OR CREDIT (000)	NET COST OR SAVINGS PER SCFM \$/SCFM
RESIDENCE TIME							
0.50SEC							
NO HEAT RECOVERY							
	700.	567.	164.	268.	0.	372.	531.80
	5000.	1039.	301.	1374.	0.	1675.	335.05
	20000.	2080.	603.	5442.	0.	6045.	302.25
	50000.	3705.	1079.	13577.	0.	14652.	293.04
	100000.	5755.	1669.	27137.	0.	28806.	288.64
RESIDENCE TIME							
0.50SEC							
HEAT RECOVERY							
	700.	585.	170.	225.	28.	367.	524.85
	5000.	1108.	321.	1389.	198.	1512.	302.41
	20000.	2471.	717.	5449.	792.	5374.	268.69
	50000.	4127.	1197.	13569.	1980.	12786.	255.72
	100000.	6618.	1919.	27102.	3960.	25061.	250.61
RESIDENCE TIME							
0.75SEC							
NO HEAT RECOVERY							
	700.	568.	165.	208.	0.	373.	532.22
	5000.	1074.	311.	1374.	0.	1685.	337.08
	20000.	2258.	655.	5442.	0.	6097.	304.83
	50000.	4022.	1166.	13577.	0.	14744.	294.88
	100000.	6321.	1833.	27137.	0.	28970.	289.70
RESIDENCE TIME							
0.75SEC							
HEAT RECOVERY							
	700.	589.	170.	225.	78.	368.	525.26
	5000.	1141.	331.	1389.	198.	1522.	304.44
	20000.	2649.	768.	5449.	792.	5455.	271.27
	50000.	4414.	1289.	13569.	1980.	12878.	257.56
	100000.	7184.	2083.	27102.	3960.	25226.	252.26

OFFGAS HEAT CONTENT 50.00 BTU/SCF
COMBUSTION TEMPERATURE 1400 F

CASE	OFFGAS FLOW	CAPITAL COST	FIXED COST	OPERATING COST-OR-CREDIT OPERATING COST	RECOVERY CREDIT	NET ANNUALIZED COST OR CREDIT	NET COST OR SAVINGS PER SCFH
	SCFH	(000)	(000)	(000)	(000)	(000)	\$/SCFH
RESIDENCE TIME							
0.50SEC	700.	587.	164.	420.	0.	585.	845.26
NO HEAT RECOVERY	5000.	1039.	301.	2891.	0.	3193.	438.51
	20000.	2080.	603.	11511.	0.	12114.	405.70
	50000.	3705.	1074.	28750.	0.	29825.	396.49
	100000.	5755.	1669.	57482.	0.	59151.	591.51
RESIDENCE TIME							
0.50SEC	700.	585.	170.	438.	28.	580.	828.20
HEAT RECOVERY	5000.	1106.	321.	2907.	189.	3029.	405.86
	20000.	2471.	717.	11518.	792.	11443.	572.15
	50000.	4127.	1197.	26742.	1980.	27959.	359.17
	100000.	6618.	1919.	57447.	3960.	53407.	554.07
RESIDENCE TIME							
0.75SEC	700.	548.	165.	420.	0.	585.	835.67
NO HEAT RECOVERY	5000.	1074.	311.	2891.	0.	3203.	410.54
	20000.	2258.	655.	11511.	0.	12166.	408.29
	50000.	4022.	1166.	28750.	0.	29917.	398.33
	100000.	6321.	1833.	57482.	0.	59315.	593.15
RESIDENCE TIME							
0.75SEC	700.	586.	170.	438.	28.	580.	828.77
HEAT RECOVERY	5000.	1141.	331.	2907.	198.	3039.	407.89
	20000.	2649.	768.	11518.	792.	11495.	374.73
	50000.	4444.	1289.	28742.	1980.	28051.	361.01
	100000.	7184.	2083.	57447.	3960.	53571.	555.71

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OFFGAS HEAT CONTENT 1.00 BTU/SCF
COMBUSTION TEMPERATURE 1600 F

CASE	OFFGAS FLOW SCFH	CAPITAL COST (000)	FIXED COST (000)	OPERATING COST-OR-CREDIT OPERATING COST (000)	RECOVERY CREDIT (000)	NET ANNUALIZED COST OR CREDIT (000)	NET COST OR SAVINGS PER SCFH \$/SCFH
RESIDENCE TIME 0.50SEC							
NO HEAT RECOVERY	700.	587.	170.	88.	0.	258.	368.24
	5000.	1108.	321.	515.	0.	835.	167.20
	20000.	2498.	782.	2005.	0.	2787.	139.36
	50000.	4107.	1191.	4985.	0.	6176.	123.52
	100000.	6377.	1849.	9952.	0.	11801.	118.01
RESIDENCE TIME 0.50SEC							
HEAT RECOVERY	700.	592.	172.	105.	38.	239.	341.68
	5000.	1190.	345.	530.	272.	603.	120.62
	20000.	2798.	811.	2010.	1086.	1735.	86.77
	50000.	4701.	1363.	4971.	2716.	3619.	72.38
	100000.	7352.	2190.	9967.	5431.	6666.	66.66
RESIDENCE TIME 0.75SEC							
NO HEAT RECOVERY	700.	594.	172.	88.	0.	260.	371.18
	5000.	1149.	333.	515.	0.	848.	167.58
	20000.	2910.	844.	2005.	0.	2849.	142.43
	50000.	4490.	1302.	4985.	0.	6287.	125.74
	100000.	7057.	2047.	9952.	0.	11996.	119.98
RESIDENCE TIME 0.75SEC							
HEAT RECOVERY	700.	599.	174.	105.	38.	241.	343.98
	5000.	1231.	357.	530.	272.	615.	122.99
	20000.	3010.	873.	2010.	1086.	1797.	89.84
	50000.	5084.	1474.	4971.	2716.	3730.	74.60
	100000.	8232.	2387.	9907.	5431.	6863.	68.63

OFFGAS HEAT CONTENT 10.00 BTU/SCF
COMBUSTION TEMPERATURE 1400 F

CASE	OFFGAS FLOW SCFM	CAPITAL COST (000)	FIXED COST (000)	OPERATING COST-OR-CREDIT (OPERATING COST (000)	RECOVERY CREDIT (000)	NET ANNUALIZED COST OR CREDIT (000)	NET COST OR SAVINGS PER SCFM 1/SCFM
RESIDENCE TIME 0.50SEC NO HEAT RECOVERY	700. 5000. 20000. 50000. 100000.	597. 1108. 2698. 4107. 6377.	170. 321. 782. 1191. 1849.	151. 970. 3825. 9537. 19055.	0. 0. 0. 0. 0.	321. 1291. 4808. 10728. 20905.	459.27 258.24 230.40 211.55 209.05
RESIDENCE TIME 0.50SEC HEAT RECOVERY	700. 5000. 20000. 50000. 100000.	592. 1190. 2798. 4701. 7552.	172. 345. 811. 1363. 2190.	169. 985. 3831. 9523. 19010.	38. 272. 1086. 2716. 5431.	362. 1058. 3556. 8171. 15769.	432.12 211.65 177.80 161.42 157.69
RESIDENCE TIME 0.75SEC NO HEAT RECOVERY	700. 5000. 20000. 50000. 100000.	594. 1149. 2910. 4490. 7057.	172. 333. 844. 1302. 2047.	151. 970. 3825. 9537. 19055.	0. 0. 0. 0. 0.	324. 1303. 4669. 10839. 21102.	462.17 260.62 233.47 216.78 211.02
RESIDENCE TIME 0.75SEC HEAT RECOVERY	700. 5000. 20000. 50000. 100000.	599. 1231. 3010. 5084. 8232.	174. 352. 873. 1474. 2387.	169. 995. 3831. 9523. 19010.	38. 272. 1086. 2716. 5431.	305. 1070. 3518. 8282. 15967.	435.02 213.03 180.88 163.64 159.67

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OFFGAS HEAT CONTENT 13.00 BTU/SCF
COMBUSTION TEMPERATURE 1800 F

CASE	OFFGAS FLOW SCFM	CAPITAL COST (000)	FIXED COST (000)	OPERATING COST-OR-CREDIT (000)	RECOVERY CREDIT (000)	NET ANNUALIZED COST OR CREDIT (000)	NET LOST OR SAVINGS PER SCFM \$/SCFM
RESIDENCE TIME 0.50SEC							
NO HEAT RECOVERY							
	700.	587.	176.	173.	0.	343.	489.62
	5000.	1108.	321.	1122.	0.	1443.	288.58
	20000.	2698.	782.	4432.	0.	5215.	260.74
	50000.	6107.	1191.	11054.	0.	12245.	244.90
	100000.	6377.	1849.	22690.	0.	23939.	239.39
RESIDENCE TIME 0.50SEC							
HEAT RECOVERY							
	700.	592.	172.	190.	38.	324.	462.46
	5000.	1190.	345.	1136.	272.	1210.	242.00
	20000.	2798.	811.	4438.	1086.	4163.	208.15
	50000.	4701.	1363.	11041.	2718.	9688.	193.74
	100000.	7552.	2196.	22645.	5431.	18604.	186.04
RESIDENCE TIME 0.75SEC							
NO HEAT RECOVERY							
	700.	594.	172.	173.	0.	345.	492.02
	5000.	1149.	333.	1122.	0.	1455.	290.96
	20000.	2910.	844.	4432.	0.	5276.	263.81
	50000.	4490.	1302.	11054.	0.	12356.	247.12
	100000.	7057.	2047.	22690.	0.	24137.	241.37
RESIDENCE TIME 0.75SEC							
HEAT RECOVERY							
	700.	599.	174.	190.	38.	326.	465.44
	5000.	1231.	357.	1136.	272.	1222.	243.38
	20000.	3010.	873.	4438.	1086.	4224.	211.22
	50000.	5084.	1474.	11041.	2718.	9799.	195.99
	100000.	6232.	2387.	22645.	5431.	19061.	190.61

OFFGAS HEAT CONTENT 20.00 BTU/SCF
COMBUSTION TEMPERATURE 1600 F

CASE	OFFGAS FLOW SCFM	CAPITAL COST (000)	FIXED COST (000)	OPERATING COST-OR-CREDIT OPERATING COST (000)	RECOVERY CREDIT (000)	NET ANNUALIZED COST OR CREDIT (000)	NET COST OR SAVINGS PER SCFM \$/SCFM
RESIDENCE TIME 0.50SEC NO HEAT RECOVERY	700. 5000. 20000. 50000. 100000.	587. 1108. 2498. 4107. 6377.	170. 321. 782. 1191. 1849.	222. 1476. 5849. 14594. 29171.	0. 0. 0. 0. 0.	392. 1797. 6631. 15785. 31026.	560.43 359.39 331.58 315.71 310.20
RESIDENCE TIME 0.50SEC HEAT RECOVERY	700. 5000. 20000. 50000. 100000.	582. 1190. 2798. 4701. 7552.	172. 345. 811. 1363. 2190.	240. 1490. 5854. 14581. 29126.	38. 272. 1086. 2716. 5431.	373. 1564. 5579. 13229. 25885.	533.27 312.80 278.96 264.57 258.85
RESIDENCE TIME 0.75SEC NO HEAT RECOVERY	700. 5000. 20000. 50000. 100000.	594. 1149. 2910. 4890. 7057.	172. 333. 844. 1302. 2047.	222. 1476. 5849. 14594. 29171.	0. 0. 0. 0. 0.	394. 1809. 6652. 15896. 31217.	563.33 361.77 334.62 317.93 312.17
RESIDENCE TIME 0.75SEC HEAT RECOVERY	700. 5000. 20000. 50000. 100000.	599. 1231. 3010. 5084. 8232.	174. 357. 873. 1474. 2387.	240. 1490. 5854. 14581. 29126.	38. 272. 1086. 2716. 5431.	375. 1576. 5641. 13310. 25082.	536.17 315.18 262.03 266.79 260.62

OFFGAS HEAT CONTENT 50.00 BTU/SCF
COMBUSTION TEMPERATURE 1500 F

CASE	OFFGAS FLOW SCFM	CAPITAL COST (000)	FIXED COST (000)	OPERATING COST (000)	RECOVERY CREDIT (000)	MEI ANNUALIZED COST OR CREDIT (000)	NET COST OR SAVINGS PER SCFM \$/SCFM
RESIDENCE TIME							
0.50SEC							
NO HEAT RECOVERY							
	700.	587.	170.	434.	0.	605.	863.88
	5000.	1108.	321.	2993.	0.	3314.	462.85
	20000.	2698.	782.	11910.	0.	12700.	435.60
	50000.	4107.	1191.	29767.	0.	30958.	419.16
	100000.	6377.	1849.	59516.	0.	61365.	413.65
RESIDENCE TIME							
0.50SEC							
HEAT RECOVERY							
	700.	592.	172.	452.	38.	586.	836.73
	5000.	1190.	345.	3008.	272.	3081.	416.26
	20000.	2798.	811.	11923.	1086.	11648.	582.41
	50000.	4701.	1363.	29754.	2716.	28401.	568.03
	100000.	7552.	2190.	59471.	5431.	56230.	562.30
RESIDENCE TIME							
0.75SEC							
NO HEAT RECOVERY							
	700.	594.	172.	434.	0.	607.	866.74
	5000.	1149.	333.	2993.	0.	3326.	465.22
	20000.	2910.	844.	11918.	0.	12762.	538.08
	50000.	4490.	1302.	29767.	0.	31069.	421.38
	100000.	7057.	2047.	59516.	0.	61563.	415.63
RESIDENCE TIME							
0.75SEC							
HEAT RECOVERY							
	700.	599.	174.	452.	38.	588.	839.43
	5000.	1231.	357.	3008.	272.	3093.	418.64
	20000.	3010.	873.	11923.	1086.	11710.	585.48
	50000.	5084.	1474.	29754.	2716.	28312.	570.25
	100000.	8232.	2387.	59471.	5431.	56427.	564.27

APPENDIX B
ANNUAL COST DATA FOR HALOGEN-CONTAINING VOC CONTROL

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OFFGAS HEAT CONTENT 1.00 BTU/SCF
COMBUSTION TEMPERATURE 1800 F

CASE	OFFGAS FLOW SCFM	CAPITAL COST (000)	FIXED COST (000)	OPERATING COST (000)	RECOVERY CREDIT (000)	NET ANNUALIZED COST OR CREDIT (000)	NET COST OR SAVINGS PER SCFM 1/SCFM
RESIDENCE TIME 0.50SEC NO HEAT RECOVERY	700.	576.	197.	160.	0.	297.	424.39
	5000.	1726.	356.	606.	0.	962.	192.30
	20000.	2612.	757.	2370.	0.	3127.	156.37
	50000.	3763.	1181.	3898.	0.	7279.	145.58
RESIDENCE TIME 0.50SEC HEAT RECOVERY	100000.	7443.	2158.	11777.	0.	13936.	139.36
	700.	713.	207.	118.	52.	272.	392.15
	5000.	1349.	391.	620.	372.	639.	127.78
	20000.	2885.	837.	2372.	1489.	1720.	85.58
RESIDENCE TIME 0.75SEC NO HEAT RECOVERY	50000.	5590.	1621.	5876.	3723.	3774.	75.48
	100000.	9298.	2392.	11715.	7446.	6966.	69.66
	700.	682.	199.	100.	0.	298.	425.25
	5000.	1087.	373.	606.	0.	979.	193.84
RESIDENCE TIME 0.75SEC HEAT RECOVERY	20000.	2773.	804.	2370.	0.	3176.	158.70
	50000.	5231.	1517.	5898.	0.	7415.	140.29
	100000.	8375.	2429.	11777.	0.	14266.	142.66
	700.	717.	208.	118.	52.	274.	390.81
RESIDENCE TIME 0.50SEC NO HEAT RECOVERY	5000.	1510.	109.	620.	372.	639.	131.21
	20000.	3047.	384.	2372.	1489.	1766.	88.70
	50000.	6058.	1757.	5876.	3723.	3910.	78.19
	100000.	10230.	2767.	11715.	7446.	7236.	72.36

OFFGAS HEAT CONTENT 10.00 BTU/SCF
COMBUSTION TEMPERATURE 1800 F

CASE	OFFGAS FLOW SCFM	CAPITAL COST (000)	FIXED COST (000)	OPERATING COST-OR-CREDIT OPERATING COST (000)	RECOVERY CREDIT (000)	NET ANNUALIZED COST OR CREDIT (000)	NET COST OR SAVINGS PER SCFM \$/SCFM
RESIDENCE TIME 0.50SEC NO HEAT RECOVERY	700. 5000. 20000. 50000. 100000.	678. 1225. 2612. 4753. 7443.	197. 356. 757. 1381. 2158.	125. 781. 3671. 7651. 15284.	0. 0. 0. 0. 0.	321. 1137. 3829. 9032. 17443.	457.25 227.27 191.43 180.65 174.53
RESIDENCE TIME 0.50SEC HEAT RECOVERY	700. 5000. 20000. 50000. 100000.	713. 1319. 2886. 5590. 9298.	207. 391. 837. 1421. 2696.	142. 795. 3073. 7629. 15222.	52. 372. 1489. 3723. 7446.	297. 814. 2421. 3527. 10473.	424.22 162.84 121.05 110.54 104.73
RESIDENCE TIME 0.75SEC NO HEAT RECOVERY	700. 5000. 20000. 50000. 100000.	682. 1287. 2773. 5231. 8375.	198. 373. 804. 1517. 2479.	125. 781. 3071. 7651. 15284.	0. 0. 0. 0. 0.	323. 1155. 3875. 9148. 17713.	460.92 230.91 193.77 183.34 177.13
RESIDENCE TIME 0.75SEC HEAT RECOVERY	700. 5000. 20000. 50000. 100000.	717. 1310. 3047. 6058. 10230.	208. 409. 884. 1757. 2967.	142. 785. 3073. 7629. 15222.	52. 372. 1489. 3723. 7446.	298. 832. 2468. 5643. 10743.	425.87 166.38 123.38 113.26 107.13

OFFGAS HEAT CONTENT 13.00 BTH/SCF
COMBUSTION TEMPERATURE 1800 F

CASE	OFFGAS FLOW SCFH	CAPITAL COST (000)	FIXED COST (000)	OPERATING COST-OR-CREDIT OPERATING COST (000)	RECOVERY CREDIT (000)	NET ANNUALIZED COST OR CREDIT (000)	NET COST OR SAVINGS PER SCFH \$/SCFH
RESIDENCE TIME 0.50SEC NO HEAT RECOVERY	700. 5000. 20000. 50000. 100000.	678. 1226. 2612. 3743. 7443.	197. 356. 757. 1381. 2158.	133. 840. 3305. 8235. 16453.	0. 0. 0. 0. 0.	330. 1195. 4062. 9417. 18611.	470.95 239.06 203.12 192.34 186.11
RESIDENCE TIME 0.50SEC HEAT RECOVERY	700. 5000. 20000. 50000. 100000.	713. 1349. 2886. 5590. 9298.	207. 391. 837. 1621. 2698.	150. 854. 3307. 8214. 16391.	52. 372. 1409. 3723. 7446.	305. 873. 2655. 6112. 13641.	435.90 174.53 132.74 122.23 116.41
RESIDENCE TIME 0.75SEC NO HEAT RECOVERY	700. 5000. 20000. 50000. 100000.	682. 1287. 2723. 5231. 8375.	198. 373. 804. 1517. 2429.	133. 840. 3305. 8235. 16453.	0. 0. 0. 0. 0.	331. 1213. 4109. 9752. 18882.	472.61 242.60 205.46 195.05 188.82
RESIDENCE TIME 0.75SEC HEAT RECOVERY	700. 5000. 20000. 50000. 100000.	717. 1410. 3047. 6058. 10230.	208. 409. 884. 1757. 2967.	156. 854. 3307. 8214. 16391.	52. 372. 1489. 3723. 7446.	306. 890. 2761. 6247. 11912.	437.56 178.07 155.07 124.95 119.12

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OFF-GAS HEAT CONTENT 20.00 BTU/SCF
COMBUSTION TEMPERATURE 1800 F

CASE	OFF-GAS FLOW SCFH	CAPITAL COST (000)	FIXED COST (000)	OPERATING COST-OR-CREDIT COST (000)	RECOVERY CREDIT (000)	NET ANNUALIZED COST OR CREDIT (000)	NET COST OR SAVINGS PER SCFH \$/SCFH
RESIDENCE TIME 0.50SEC NO HEAT RECOVERY	700. 5000. 20000. 50000. 100000.	678. 1226. 2612. 3763. 7443.	197. 356. 757. 1381. 2158.	152. 976. 3850. 9599. 19180.	0. 0. 0. 0. 0.	349. 1332. 4608. 10980. 21339.	498.22 264.33 230.40 219.61 213.39
RESIDENCE TIME 0.50SEC HEAT RECOVERY	700. 5000. 20000. 50000. 100000.	713. 1349. 2886. 5590. 9298.	207. 391. 837. 1621. 2696.	170. 990. 3853. 9577. 19119.	52. 372. 1489. 3723. 7446.	324. 1009. 3200. 7175. 14369.	463.18 201.81 140.01 149.51 143.69
RESIDENCE TIME 0.75SEC NO HEAT RECOVERY	700. 5000. 20000. 50000. 100000.	682. 1287. 2773. 5231. 8375.	198. 373. 804. 1517. 2429.	152. 976. 3850. 9599. 19180.	0. 0. 0. 0. 0.	350. 1349. 4655. 11116. 21609.	495.83 269.87 232.73 222.32 216.09
RESIDENCE TIME 0.75SEC HEAT RECOVERY	700. 5000. 20000. 50000. 100000.	717. 1410. 3047. 4058. 10230.	208. 409. 884. 1757. 2767.	170. 990. 3853. 9577. 19119.	52. 372. 1489. 3723. 7446.	325. 1027. 3247. 7611. 14639.	465.84 205.35 162.35 152.22 146.39

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OF GAS HEAT CONTENT 50.00 BTU/SCF
COMBUSTION TEMPERATURE 1800 F

CASE	OFFGAS FLOW	CAPITAL COST	FIXED COST	OPERATING COST	RECOVERY CREDIT	NET ANNUALIZED COST OR CREDIT	NET COST OF SAVINGS PER SCFM \$/SCFH
	SCFH	(000)	(000)	(000)	(000)	(000)	
RESIDENCE TIME 0.50SEC NO HEAT RECOVERY	700. 5000. 20000. 50000. 100000.	678. 1226. 2612. 4763. 7443.	197. 356. 757. 1381. 2158.	234. 1561. 6168. 15444. 30870.	0. 0. 0. 0. 0.	431. 1916. 6946. 16875. 33628.	615.11 383.22 347.29 336.50 330.28
RESIDENCE TIME 0.50SEC HEAT RECOVERY	700. 5000. 20000. 50000. 100000.	713. 1349. 2886. 5590. 9298.	207. 391. 837. 1621. 2696.	251. 1575. 6190. 15422. 30808.	52. 372. 1489. 3723. 7446.	406. 1593. 5538. 13320. 26058.	580.07 318.70 276.90 266.40 260.58
RESIDENCE TIME 0.75SEC NO HEAT RECOVERY	700. 5000. 20000. 50000. 100000.	682. 1287. 2773. 5231. 8375.	198. 373. 804. 1517. 2429.	234. 1561. 6188. 15444. 30870.	0. 0. 0. 0. 0.	432. 1934. 6992. 16961. 33298.	616.77 386.76 349.62 339.21 332.98
RESIDENCE TIME 0.75SEC HEAT RECOVERY	700. 5000. 20000. 50000. 100000.	717. 1310. 3047. 6058. 10230.	208. 409. 884. 1757. 2967.	251. 1575. 6190. 15422. 30808.	52. 372. 1489. 3723. 7446.	407. 1611. 5585. 14356. 26328.	581.73 322.24 279.24 269.11 263.28

OFFGAS HEAT CAPACITY 1000 BTU/SCF
COMBUSTION TEMPERATURE 1800 F

CASE	OFFGAS FLOW SCFM	CAPITAL COST (000)	FIXED COST (000)	OPERATING COST-OR-CREDIT OPERATING COST (000)	RECOVERY CREDIT (000)	NET ANNUALIZED COST OR CREDIT (000)	NET COST OF SAMPLE PER SCFM \$/SCFM
RESIDENCE TIME							
0.50SEC							
NO HEAT RECOVERY							
	700.	578.	197.	370.	0.	567.	802.22
	5000.	1226.	356.	2535.	0.	2890.	578.04
	20000.	2612.	757.	10085.	0.	10842.	542.11
	50000.	4763.	1481.	23185.	0.	26366.	531.32
	100000.	7443.	2158.	56351.	0.	52510.	525.10
RESIDENCE TIME							
0.50SEC							
HEAT RECOVERY							
	700.	713.	267.	388.	52.	542.	773.89
	5000.	1349.	391.	2559.	372.	2568.	513.53
	20000.	2884.	837.	10087.	1489.	9434.	471.72
	50000.	5590.	1421.	25183.	3723.	23061.	461.22
	100000.	9298.	2694.	50289.	7446.	45540.	455.40
RESIDENCE TIME							
0.75SEC							
NO HEAT RECOVERY							
	700.	552.	198.	370.	0.	568.	811.59
	5000.	1287.	373.	2535.	0.	2908.	581.56
	20000.	2773.	804.	10065.	0.	10889.	544.44
	50000.	5231.	1517.	25185.	0.	26702.	534.03
	100000.	8375.	2429.	56351.	0.	52780.	527.80
RESIDENCE TIME							
0.75SEC							
HEAT RECOVERY							
	700.	717.	208.	388.	52.	544.	776.35
	5000.	1310.	509.	2549.	372.	2585.	517.05
	20000.	3047.	884.	10087.	1489.	9481.	474.55
	50000.	4058.	1757.	25183.	3723.	23197.	463.93
	100000.	10230.	2967.	56289.	7446.	45816.	458.16

OFFGAS HEAT CONTENT 1,000 BTU/SCF
COMBUSTION TEMPERATURE 2,000 F

CASE	OFFGAS FLOW SCFM	CAPITAL COST (000)	FIXED COST (000)	OPERATING COST-OR-CREDIT OPERATING COST (000)	RECOVERY CREDIT (000)	NET ANNUALIZED COST OR CREDIT (000)	NET COST OR SAVING PER SCFM 1/SCFM
RESIDENCE TIME 0.50SEC NO HEAT RECOVERY	700. 5000. 20000. 50000.	736. 1446. 3458. 6104.	213. 419. 1003. 1857.	146. 929. 3663. 9130.	0. 0. 0. 0.	359. 1349. 4666. 10977.	512.73 269.73 233.28 219.74
RESIDENCE TIME 0.50SEC HEAT RECOVERY	700. 5000. 20000. 50000.	781. 1792. 4818. 8618.	226. 520. 1397. 2441.	163. 940. 3653. 9079.	86. 613. 2453. 6132.	303. 847. 2508. 5368.	433.14 169.38 129.89 107.77
RESIDENCE TIME 0.75SEC NO HEAT RECOVERY	700. 5000. 20000. 50000.	753. 1557. 3712. 7207.	218. 452. 1076. 2090.	146. 929. 3663. 9130.	0. 0. 0. 0.	364. 1381. 4739. 11220.	519.82 276.17 236.96 224.40
RESIDENCE TIME 0.75SEC HEAT RECOVERY	700. 5000. 20000. 50000.	798. 1903. 5072. 9221.	231. 552. 1471. 2674.	163. 940. 3653. 9079.	86. 613. 2453. 6132.	308. 879. 2671. 5621.	440.17 175.82 133.57 112.42

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OFFGAS HEAT CONTENT 10.00 BTU/SCF
COMBUSTION TEMPERATURE 2200 F

CASE	OFFGAS FLOW SCFM	CAPITAL COST (000)	FIXED COST (000)	OPERATING COST-ON-CREDIT OPERATING COST (000)	RECOVERY CREDIT (000)	NET ANNUALIZED COST OR CREDIT (000)	NET COST OR SAVINGS PER SCFM \$/SCFM
RESIDENCE TIME 0.50SEC NO HEAT RECOVERY	700. 5000. 20000. 50000.	734. 1440. 3438. 8363.	213. 418. 997. 1845.	168. 1090. 4305. 10735.	0. 0. 0. 0.	381. 1507. 5302. 12081.	544.08 301.45 265.09 251.61
RESIDENCE TIME 0.50SEC HEAT RECOVERY	700. 5000. 20000. 50000.	779. 1787. 4801. 8387.	226. 518. 1392. 2432.	185. 1101. 4295. 10684.	86. 413. 2452. 4132.	325. 1006. 3235. 6985.	464.53 201.18 161.75 139.70
RESIDENCE TIME 0.75SEC NO HEAT RECOVERY	700. 5000. 20000. 50000.	751. 1531. 3692. 7166.	218. 450. 1071. 2078.	168. 1090. 4305. 10735.	0. 0. 0. 0.	386. 1539. 5376. 12814.	551.10 307.89 268.78 256.27
RESIDENCE TIME 0.75SEC HEAT RECOVERY	700. 5000. 20000. 50000.	796. 1898. 5055. 9190.	231. 550. 1466. 2665.	185. 1101. 4295. 10684.	86. 413. 2452. 4132.	330. 1038. 3309. 7218.	471.57 207.62 145.43 141.35

OFFGAS HEAT CONTENT 14.00 BTU/SCF
COMBUSTION TEMPERATURE 2200 F

CASE	OFFGAS FLOW SCFH	CAPITAL COST (000)	FIXED COST (000)	OPERATING COST-OR-CREDIT COST (000)	RECOVERY CREDIT (000)	NET ANNUALIZED COST OR CREDIT (000)	NET COST OR SAVINGS PER SCFH \$/SCFH
RESIDENCE TIME 0.50SEC NO HEAT RECOVERY	700. 5000. 20000. 50000.	733. 1437. 3431. 6350.	213. 417. 995. 1841.	176. 1143. 4519. 11271.	0. 0. 0. 0.	388. 1560. 5514. 13112.	554.48 312.03 275.70 262.24
RESIDENCE TIME 0.50SEC HEAT RECOVERY	700. 5000. 20000. 50000.	79. 1785. 4775. 8377.	226. 518. 1391. 2429.	193. 1154. 4568. 11220.	86. 613. 2453. 6132.	332. 1059. 3447. 7917.	475.00 211.78 172.37 150.34
RESIDENCE TIME 0.75SEC NO HEAT RECOVERY	700. 5000. 20000. 50000.	750. 1518. 3685. 7153.	218. 449. 1049. 2074.	176. 1143. 4519. 11271.	0. 0. 0. 0.	393. 1592. 5588. 13345.	561.53 318.46 279.38 266.90
RESIDENCE TIME 0.75SEC HEAT RECOVERY	700. 5000. 20000. 50000.	796. 1896. 5049. 9180.	231. 550. 1464. 2662.	193. 1154. 4569. 11220.	86. 613. 2453. 6132.	337. 1091. 3521. 7750.	482.64 218.22 176.05 155.00

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OFFGAS HEAT CONTENT 20.00 BTU/SCF
COMBUSTION TEMPERATURE 2200 F

CASE	OFFGAS FLOW SCFH	CAPITAL COST (000)	FIXED COST (000)	OPERATING COST-OR-CREDIT OPERATING COST (000)	RECOVERY CREDIT (000)	NET ANNUALIZED COST OR CREDIT (000)	NET COST OR SAVINGS PER SCFH \$/SCFH
RESIDENCE TIME 0.50SEC NO HEAT RECOVERY	700. 5000. 20000. 50000.	732. 1832. 3415. 6318.	212. 415. 990. 1832.	193. 1268. 5018. 12319.	0. 0. 0. 0.	405. 1483. 6009. 14351.	578.81 336.70 300.44 287.03
RESIDENCE TIME 0.50SEC HEAT RECOVERY	700. 5000. 20000. 50000.	777. 1781. 4782. 8354.	225. 517. 1387. 2423.	210. 1279. 5069. 12468.	86. 613. 2453. 6132.	356. 1183. 3943. 8759.	409.42 236.51 197.14 175.17
RESIDENCE TIME 0.75SEC NO HEAT RECOVERY	706. 5000. 20000. 50000.	749. 1513. 3569. 7121.	217. 448. 1064. 2065.	193. 1268. 5018. 12519.	0. 0. 0. 0.	410. 1716. 6083. 14584.	585.65 331.14 304.13 291.68
RESIDENCE TIME 0.75SEC HEAT RECOVERY	706. 5000. 20000. 50000.	794. 1892. 5036. 9157.	230. 549. 1461. 2655.	210. 1279. 5009. 12468.	86. 613. 2453. 6132.	355. 1215. 4017. 8991.	506.44 242.95 200.43 179.83

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OFFGAS HEAT CONTENT 56.00 Btu/SCF
COMBUSTION TEMPERATURE 2200 F

CASE	OFFGAS FLOW SCFM	CAPITAL COST (000)	FIXED COST (000)	OPERATING COST-OR-CREDIT OPERATING COST (000)	RECOVERY CREDIT (000)	NET ANNUALIZED COST OR CREDIT (000)	NET COST OR SAVINGS PER SCFM \$/SCFM
RESIDENCE TIME 0.50SEC NO HEAT RECOVERY	700. 5000. 20000. 50000.	725. 1410. 3348. 6183.	210. 409. 971. 1793.	268. 1803. 7159. 17870.	0. 0. 0. 0.	478. 2212. 8130. 19663.	683.07 432.44 406.49 393.26
RESIDENCE TIME 0.50SEC HEAT RECOVERY	700. 5000. 20000. 50000.	772. 1763. 4726. 8252.	224. 511. 1376. 2393.	285. 1814. 7149. 17819.	86. 413. 2453. 6132.	423. 1212. 6067. 14080.	604.09 342.50 363.84 281.60
RESIDENCE TIME 0.75SEC NO HEAT RECOVERY	700. 5000. 20000. 50000.	742. 1521. 3502. 6986.	215. 441. 1045. 2026.	268. 1803. 7159. 17870.	0. 0. 0. 0.	483. 2214. 8263. 19896.	490.11 448.88 416.17 397.92
RESIDENCE TIME 0.75SEC HEAT RECOVERY	700. 5000. 20000. 50000.	789. 1874. 4980. 9055.	229. 544. 1441. 2426.	285. 1814. 7149. 17819.	86. 413. 2453. 6132.	428. 1745. 6146. 14313.	611.13 348.94 307.02 284.26

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OFFGAS HEAT CONTENT 160.00 BTU/SCF
COMBUSTION TEMPERATURE 2200 F

CASE	OFFGAS FLOW SCFH	CAPITAL COST (000)	FIXED COST (000)	OPERATING COST-OK-CREDIT OPERATING LOST (000)	RECOVERY CREDIT (000)	NET ANNUALIZED COST OR CREDIT (000)	NET UNIT OR SAVING PER SCFH \$/SCFH
RESIDENCE TIME 0.50SEC NO HEAT RECOVERY	700. 5000. 20000. 50000.	714. 1374. 3236. 5959.	207. 398. 939. 1728.	393. 2695. 10726. 24788.	0. 0. 0. 0.	600. 3093. 11665. 28316.	854.83 618.69 583.23 570.33
RESIDENCE TIME 0.50SEC HEAT RECOVERY	700. 5000. 20000. 50000.	762. 1734. 4631. 8084.	221. 503. 1343. 2344.	410. 2706. 10716. 24737.	84. 413. 2453. 6132.	545. 2596. 9607. 22949.	778.54 519.15 480.34 458.99
RESIDENCE TIME 0.75SEC NO HEAT RECOVERY	700. 5000. 20000. 50000.	731. 1485. 3490. 6762.	217. 431. 1012. 1941.	393. 2695. 10726. 24788.	0. 0. 0. 0.	605. 3126. 11738. 28749.	863.88 623.12 584.91 574.98
RESIDENCE TIME 0.75SEC HEAT RECOVERY	700. 5000. 20000. 50000.	779. 1835. 4885. 8887.	226. 535. 1417. 2477.	410. 2706. 10716. 24737.	86. 413. 2453. 6132.	550. 2628. 9680. 23182.	785.58 525.58 481.02 461.65

OFFGAS HEAT CONTENT 1.00 BTU/SCF
COMBUSTION TEMPERATURE 2600 F

CASE	OFFGAS FLOW SCFM	CAPITAL COST (000)	FIXED COST (000)	OPERATING COST-OR-CREDIT OPERATING COST (000)	RECOVERY CREDIT (000)	NET ANNUALIZED COST OR CREDIT (000)	NET COST OR SAVINGS PER SCFM \$/SCFM
RESIDENCE TIME 0.50SEC NO HEAT RECOVERY	700. 5000. 20000.	1190. 2649. 5807.	345. 748. 1684.	270. 1821. 7230.	0. 0. 0.	615. 2589. 8914.	879.20 517.85 445.71
RESIDENCE TIME 0.50SEC HEAT RECOVERY	700. 5000. 20000.	1208. 2903. 7238.	350. 842. 2099.	286. 1822. 7181.	170. 1218. 4871.	466. 1447. 4409.	665.51 289.31 220.47
RESIDENCE TIME 0.75SEC NO HEAT RECOVERY	700. 5000. 20000.	1240. 2879. 6467.	360. 835. 1875.	270. 1821. 7230.	0. 0. 0.	630. 2656. 9106.	899.92 531.19 455.28
RESIDENCE TIME 0.75SEC HEAT RECOVERY	700. 5000. 20000.	1258. 3133. 7898.	365. 909. 2290.	286. 1822. 7181.	170. 1218. 4871.	480. 1513. 4601.	686.23 302.65 230.04

OFFGAS HEAT CONTENT 10.00 BTU/SCF
COMBUSTION TEMPERATURE 2600 F

CASE	OFFGAS FLOW SCFH	CAPITAL COST (000)	FIXED COST (000)	OPERATING COST-OR-CREDIT OPERATING COST (000)	RECOVERY CREDIT (000)	NET ANNUALIZED COST OR CREDIT (000)	NET COST OR SAVINGS PER SCFH \$/SCFH
RESIDENCE TIME 0.50SEC NO HEAT RECOVERY	700. 5000. 20000.	1187. 2638. 5777.	344. 765. 1875.	292. 1973. 7836.	0. 0. 0.	636. 2738. 9512.	908.39 547.55 475.60
RESIDENCE TIME 0.50SEC HEAT RECOVERY	700. 5000. 20000.	1206. 2896. 7216.	350. 840. 2093.	307. 1974. 7787.	170. 1218. 4871.	486. 1596. 5009.	694.97 319.22 250.47
RESIDENCE TIME 0.75SEC NO HEAT RECOVERY	700. 5000. 20000.	1237. 2868. 6437.	359. 832. 1867.	292. 1973. 7836.	0. 0. 0.	650. 2804. 9703.	929.11 560.89 485.17
RESIDENCE TIME 0.75SEC HEAT RECOVERY	700. 5000. 20000.	1256. 3126. 7876.	366. 907. 2284.	307. 1974. 7787.	170. 1218. 4871.	501. 1663. 5201.	715.69 332.56 260.04

OFFGAS HEAT CONTENT 13.00 BTU/SCF
COMBUSTION TEMPERATURE 2600 F

CASE	OFFGAS FLOW SCFM	CAPITAL COST (000)	FIXED COST (000)	OPERATING COST-OR-CREDIT OPERATING COST (000)	RECOVERY CREDIT (000)	NET ANNUALIZED COST OR CREDIT (000)	NET COST OR SAVINGS PER SCFM \$/SCFM
RESIDENCE TIME 0.50SEC NO HEAT RECOVERY	700. 5000. 20000.	1186. 2635. 5768.	344. 764. 1673.	299. 2023. 8038.	0. 0. 0.	643. 2787. 9711.	918.12 557.46 485.56
RESIDENCE TIME 0.50SEC HEAT RECOVERY	700. 5000. 20000.	1205. 2894. 7209.	349. 839. 2091.	314. 2024. 7989.	170. 1218. 4871.	493. 1446. 5209.	704.79 329.20 260.47
RESIDENCE TIME 0.75SEC NO HEAT RECOVERY	700. 5000. 20000.	1236. 2965. 6428.	358. 831. 1864.	299. 2023. 8038.	0. 0. 0.	657. 2854. 9903.	938.84 570.80 495.13
RESIDENCE TIME 0.75SEC HEAT RECOVERY	700. 5000. 20000.	1255. 3124. 7869.	364. 906. 2282.	314. 2024. 7989.	170. 1218. 4871.	508. 1713. 5401.	725.51 342.54 270.04

200

OFFGAS HEAT CONTENT 20.00 BTU/SCF
COMBUSTION TEMPERATURE 2400 F

CASE	OFFGAS FLOW SCFM	CAPITAL COST (000)	FIXED COST (000)	OPERATING COST-OR-CREDIT COST (000)	RECOVERY CREDIT (000)	NET ANNUALIZED COST OR CREDIT (000)	NET COST OR SAVINGS PER SCFM \$/SCFM
RESIDENCE TIME 0.50SEC NO HEAT RECOVERY	700. 5000. 20000.	1184. 2627. 5745.	343. 762. 1666.	315. 2141. 8510.	0. 0. 0.	659. 2903. 10176.	940.83 580.56 508.80
RESIDENCE TIME 0.50SEC HEAT RECOVERY	700. 5000. 20000.	1203. 2889. 7193.	349. 838. 2086.	331. 2142. 8461.	170. 1218. 4871.	509. 1762. 5676.	727.70 352.46 283.80
RESIDENCE TIME 0.75SEC NO HEAT RECOVERY	700. 5000. 20000.	1234. 2857. 6405.	358. 829. 1857.	315. 2141. 8510.	0. 0. 0.	673. 2970. 10367.	941.54 593.90 518.37
RESIDENCE TIME 0.75SEC HEAT RECOVERY	700. 5000. 20000.	1253. 3119. 7853.	363. 904. 2277.	331. 2142. 8461.	170. 1218. 4871.	524. 1829. 5867.	748.42 365.80 293.37

OFFGAS HEAT CONTENT 50.00 BTU/SCF
COMBUSTION TEMPERATURE 2600 F

CASE	OFFGAS FLOW SCFM	CAPITAL COST (000)	FIXED COST (000)	OPERATING COST-OR-CREDIT OPERATING COST (000)	RECOVERY CREDIT (000)	NET ANNUALIZED COST OR CREDIT (000)	NET COST OR SAVINGS PER SCFM \$/SCFM
RESIDENCE TIME 0.50SEC NO HEAT RECOVERY	700. 5000. 20000.	1175. 2592. 5447.	341. 752. 1638.	386. 2646. 10530.	0. 0. 0.	727. 3398. 12168.	1038.12 679.59 608.41
RESIDENCE TIME 0.50SEC HEAT RECOVERY	700. 5000. 20000.	1197. 2866. 7122.	347. 831. 2065.	402. 2647. 10481.	170. 1218. 4871.	578. 2261. 7676.	825.89 452.17 383.80
RESIDENCE TIME 0.75SEC NO HEAT RECOVERY	700. 5000. 20000.	1225. 2822. 6307.	355. 819. 1829.	386. 2646. 10530.	0. 0. 0.	741. 3465. 12360.	1058.84 692.93 617.98
RESIDENCE TIME 0.75SEC HEAT RECOVERY	700. 5000. 20000.	1247. 3096. 7782.	362. 898. 2257.	402. 2647. 10481.	170. 1218. 4871.	593. 2328. 7867.	846.61 465.51 393.37

232

OFFGAS HEAT CONTENT 100.00 BTU/SCF
COMBUSTION TEMPERATURE 2600 F

CASE	OFFGAS FLOW SCFM	CAPITAL COST (000)	FIXED COST (000)	OPERATING COST-OR-CREDIT OPERATING COST (000)	RECOVERY CREDIT (000)	NET ANNUALIZED COST OR CREDIT (000)	NET COST OR SAVINGS PER SCFM \$/SCFM
RESIDENCE TIME 0.50SEC NO HEAT RECOVERY	700. 5000. 20000.	1160. 2535. 5485.	336. 735. 1591.	504. 3488. 13898.	0. 0. 0.	840. 4223. 15489.	1200.29 844.63 774.43
RESIDENCE TIME 0.50SEC HEAT RECOVERY	700. 5000. 20000.	1185. 2828. 7004.	344. 820. 2031.	519. 3489. 13849.	170. 1218. 4871.	693. 3092. 11009.	989.55 618.36 550.47
RESIDENCE TIME 0.75SEC NO HEAT RECOVERY	700. 5000. 20000.	1210. 2765. 6145.	351. 802. 1782.	504. 3488. 13898.	0. 0. 0.	855. 4290. 15680.	1221.00 852.97 784.00
RESIDENCE TIME 0.75SEC HEAT RECOVERY	700. 5000. 20000.	1235. 3058. 7664.	358. 887. 2223.	519. 3489. 13849.	170. 1218. 4871.	707. 3158. 11201.	1010.26 631.70 560.04

OFFGAS HEAT CONTENT 1.00 BTU/SCF
COMBUSTION TEMPERATURE 3000 F

CASE	OFFGAS FLOW SCFM	CAPITAL COST (000)	FIXED COST (000)	OPERATING COST-OR-CREDIT OPERATING COST (000)	RECOVERY CREDIT (000)	NET ANNUALIZED COST OR CREDIT (000)	NET COST OR SAVINGS PER SCFM \$/SCFM
RESIDENCE TIME 0.50SEC NO HEAT RECOVERY	700. 5000.	1870. 4878.	542. 1415.	782. 5472.	0. 0.	1324. 6887.	1891.13 1377.41
RESIDENCE TIME 0.50SEC HEAT RECOVERY	700. 5000.	2090. 5548.	606. 1609.	798. 5480.	530. 3784.	874. 3304.	1249.04 660.87
RESIDENCE TIME 0.75SEC NO HEAT RECOVERY	700. 5000.	1970. 5638.	571. 1635.	782. 5472.	0. 0.	1353. 7107.	1932.55 1421.49
RESIDENCE TIME 0.75SEC HEAT RECOVERY	700. 5000.	2190. 6308.	635. 1829.	798. 5480.	530. 3784.	903. 3525.	1290.47 704.95

OFFGAS HEAT CONTENT 10.00 BTU/SCF
COMBUSTION TEMPERATURE 3000 F

CASE	OFFGAS FLOW SCFM	CAPITAL COST (000)	FIXED COST (000)	OPERATING COST-OR-CREDIT OPERATING COST (000)	RECOVERY CREDIT (000)	NET ANNUALIZED COST OR CREDIT (000)	NET COST OR SAVINGS PER SCFM \$/SCFM
RESIDENCE TIME 0.50SEC NO HEAT RECOVERY	700. 5000.	1865. 4859.	541. 1409.	801. 5610.	0. 0.	1342. 7019.	1916.83 1403.78
RESIDENCE TIME 0.50SEC HEAT RECOVERY	700. 5000.	2086. 5537.	605. 1606.	817. 5617.	530. 3784.	893. 3438.	1275.19 687.69
RESIDENCE TIME 0.75SEC NO HEAT RECOVERY	700. 5000.	1945. 5619.	570. 1629.	801. 5610.	0. 0.	1371. 7239.	1958.26 1447.86
RESIDENCE TIME 0.75SEC HEAT RECOVERY	700. 5000.	2186. 6297.	634. 1826.	817. 5617.	530. 3784.	922. 3659.	1316.62 731.77

OFFGAS HEAT CONTENT 13.00 BTU/SCF
COMBUSTION TEMPERATURE 3000 F

CASE	OFFGAS FLOW SCFM	CAPITAL COST (000)	FIXED COST (000)	OPERATING COST-OR-CREDIT OPERATING COST (000)	RECOVERY CREDIT (000)	NET ANNUALIZED COST OR CREDIT (000)	NET COST OR SAVINGS PER SCFM \$/SCFM
RESIDENCE TIME 0.50SEC NO HEAT RECOVERY	700. 5000.	1864. 4853.	541. 1407.	807. 5656.	0. 0.	1348. 7063.	1925.40 1412.57
RESIDENCE TIME 0.50SEC HEAT RECOVERY	700. 5000.	2085. 5533.	605. 1605.	824. 5663.	530. 3784.	899. 3483.	1283.91 696.63
RESIDENCE TIME 0.75SEC NO HEAT RECOVERY	700. 5000.	1964. 5613.	570. 1628.	807. 5656.	0. 0.	1377. 7283.	1966.82 1456.65
RESIDENCE TIME 0.75SEC HEAT RECOVERY	700. 5000.	2185. 6293.	634. 1825.	824. 5663.	530. 3784.	929. 3704.	1325.33 740.71

OFFGAS HEAT CONTENT 20.00 BTU/SCF
COMBUSTION TEMPERATURE 3000 F

CASE	OFFGAS FLOW SCFM	CAPITAL COST (000)	FIXED COST (000)	OPERATING COST-OR-CREDIT OPERATING COST (000)	RECOVERY CREDIT (000)	NET ANNUALIZED COST OR CREDIT (000)	NET COST OR SAVINGS PER SCFM \$/SCFM
RESIDENCE TIME 0.50SEC NO HEAT RECOVERY	700. 5000.	1861. 4838.	540. 1403.	822. 5762.	0. 0.	1362. 7165.	1945.39 1433.08
RESIDENCE TIME 0.50SEC HEAT RECOVERY	700. 5000.	2083. 5524.	604. 1602.	839. 5770.	530. 3784.	913. 3587.	1304.24 717.49
RESIDENCE TIME 0.75SEC NO HEAT RECOVERY	700. 5000.	1961. 5598.	569. 1623.	822. 5762.	0. 0.	1391. 7386.	1986.81 1477.16
RESIDENCE TIME 0.75SEC HEAT RECOVERY	700. 5000.	2183. 6284.	633. 1822.	839. 5770.	530. 3784.	942. 3808.	1345.67 761.57

OFFGAS HEAT CONTENT 50.00 BTU/SCF
COMBUSTION TEMPERATURE 3000 F

CASE	OFFGAS FLOW SCFM	CAPITAL COST (000)	FIXED COST (000)	OPERATING COST-OR-CREDIT COST (000)	RECOVERY CREDIT (000)	NET ANNUALIZED COST OR CREDIT (000)	NET COST OR SAVINGS PER SCFM \$/SCFM
RESIDENCE TIME 0.50SEC NO HEAT RECOVERY	700. 5000.	1846. 4775.	535. 1385.	886. 6220.	0. 0.	1422. 7605.	2031.06 1520.99
RESIDENCE TIME 0.50SEC HEAT RECOVERY	700. 5000.	2072. 5487.	601. 1591.	903. 6228.	530. 3784.	974. 4034.	1391.41 806.88
RESIDENCE TIME 0.75SEC NO HEAT RECOVERY	700. 5000.	1946. 5535.	564. 1605.	886. 6220.	0. 0.	1451. 7625.	2072.49 1565.07
RESIDENCE TIME 0.75SEC HEAT RECOVERY	700. 5000.	2172. 6247.	630. 1811.	903. 6228.	530. 3784.	1003. 4255.	1432.84 850.96

OFFGAS HEAT CONTENT 100.00 BTU/SCF
COMBUSTION TEMPERATURE 3000 F

CASE	OFFGAS FLOW SCFM	CAPITAL COST (000)	FIXED COST (000)	OPERATING COST-OR-CREDIT COST (000)	RECOVERY CREDIT (000)	NET ANNUALIZED COST OR CREDIT (000)	NET COST OR SAVINGS PER SCFM \$/SCFM
RESIDENCE TIME 0.50SEC NO HEAT RECOVERY	700. 5000.	1823. 4670.	529. 1354.	993. 6983.	0. 0.	1522. 8337.	2173.85 1667.50
RESIDENCE TIME 0.50SEC HEAT RECOVERY	700. 5000.	2055. 5424.	596. 1573.	1010. 6991.	530. 3784.	1076. 4779.	1536.69 955.86
RESIDENCE TIME 0.75SEC NO HEAT RECOVERY	700. 5000.	1923. 5430.	558. 1575.	993. 6983.	0. 0.	1551. 8558.	2215.28 1711.58
RESIDENCE TIME 0.75SEC HEAT RECOVERY	700. 5000.	2155. 6184.	625. 1793.	1010. 6991.	530. 3784.	1105. 5000.	1578.11 999.94

REPORT 3

CONTROL DEVICE EVALUATION
CATALYTIC OXIDATION

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October 1980

D941

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I. INTRODUCTION

Catalytic oxidation is a control technology, currently used by many industries, that involves the oxidation of volatile organic compounds (VOC) in the presence of a catalyst. The principles and the equipment, except for the catalyst, are similar to those used for thermal oxidation, the subject of a previous control-device evaluation report.^{1*} A waste-gas stream and oxygen, usually from air, are contacted with the catalyst at a temperature that allows the oxidation reactions to proceed rapidly. The primary difference between the equipment used for catalytic oxidation and that used for thermal oxidation is the added provision for the catalyst, which is usually composed of a noble-metal coating on activated alumina. Nonprecious-metal catalysts may be used when selectivity is not critical. Because of the catalyst the oxidation reaction takes place at a lower temperature, and therefore less fuel is required to heat the waste gas than for thermal oxidation (see Table I-1 for catalytic ignition temperatures of some common VOC as given by Oxy-Catalyst, Inc., in their brochure²). The actual temperatures required will vary with different catalysts and different supports.³

The energy in the hot flue gas leaving the catalyst may be recovered by use of a recuperative heat exchanger to preheat the waste gas and combustion air or by use of a waste-heat boiler to produce steam, in the same that way energy is recovered from the hot flue gases of a thermal oxidizer. When the concentration of VOC is high enough, a catalytic oxidizer with a recuperative heat exchanger can be designed that requires no additional fuel after operating temperatures are reached.²⁻⁵

Catalytic oxidation accomplishes the same results as thermal oxidation, i.e., the oxidation of the VOC in the waste gas to water and carbon dioxide. The catalyst increases the rate of oxidation, and thus the reaction proceeds to equilibrium at a lower temperature (energy level). The reactions of the individual molecules take place at active sites on the surface of the catalyst. The VOC and oxygen are first transferred to the surface of the catalyst by diffusion in the waste gas and are then chemisorbed in the pores of the catalyst to the active sites, where the reaction (oxidation) takes place. The reaction

*See Sect. VI for the references cited in this report.

Table I-1. Catalytic Ignition Temperature for 90% Conversion*

Component	Temperature (°F)
Hydrogen	220
Acetylene	395
Carbon monoxide	425
Propyne	460
Propadiene	480
Propylene	500
Ethylene	550
<u>n</u> -Neptane	575
Benzene	575
Toluene	575
Xylene	575
Ethanol	600
Methyl ethyl ketone	700
Methyl isobutyl ketone	700
Propane	770
Ethyl acetate	775
Dimethyl formamide	800
Ethane	810
Cyclopropane	850
Methane	920

* See ref 2.

products are then desorbed from the catalyst sites and transferred by diffusion into the waste-gas stream leaving the catalyst. The heat of combustion released by the catalytic oxidation reaction is the same as that released by thermal oxidation.²⁻⁴

An advantage of catalytic oxidation over thermal oxidation is that less NO_x is formed as the results of the lower temperature used and of operating close to the required stoichiometric amount of oxygen. Because of the theoretical relationships between VOC and NO_x , attaining the ozone standard may be served better by a lower destruction of VOC with less NO_x formation.⁶

Catalytic oxidation has some limitations that do not apply to thermal oxidation. Normally the waste gas should not contain materials that poison the catalyst, such as phosphorus, bismuth, lead, arsenic, antimony, mercury, iron oxide, tin, silicon, zinc, sulfur, or halogens. Care must be taken that liquid or solid particles do not deposit on the catalyst and form a coating. In some applications it may be possible to adequately remove the poison or particulate materials, or catalytic oxidation systems may be available that will effectively handle some of the poisons, such as sulfur compounds, or some halogenated compounds. The VOC content of the waste gas should be relatively constant and low enough (or the waste gas diluted with air) so that the catalyst is not overheated and its activity destroyed. Because of safety considerations it is general practice to keep the concentration of VOC at less than 25 to 30% of the lower explosive limit. This concentration is in the range of 12 to 14 Btu/scf in air or 20 to 22 Btu/scf in nitrogen. Design for a specific application requires good basic data and experience; a pilot study is often necessary. This report is intended for use in preliminary screening studies to indicate whether additional investigation of catalytic oxidation is advisable.²⁻⁷

Catalytic oxidizers are being used successfully as emission control devices on the off-gases (waste gases) from the production of ethylene oxide, cumene, caprolactam, phthalic anhydride, bisphenol A, formaldehyde, acrylonitrile, and ethylene dichloride, and have been considered for use on maleic anhydride waste gas, all in the synthetic organic chemical manufacturing industry (SOCMI).^{2-4,7-12}

II. CATALYTIC OXIDATION SYSTEMS AND FACTORS INFLUENCING PERFORMANCE AND DESIGN

In this section the main elements of a catalytic oxidizer system are discussed (or reference is made to the thermal oxidizer report,¹ in which the elements discussed are the same). The factors influencing catalytic oxidizer design and performance are analyzed, with the analysis directed toward development of a design for a typical or base-case system. This base-case catalytic oxidizer and variations of it form the basis for the cost estimation given in Sect. IV.

A. SYSTEM DESCRIPTION

The catalyst bed in a catalytic oxidizer usually follows a burner and mixing chamber (see Fig. II-1). The waste gas enters the mixing chamber and is heated by mixing with the combustion products from the burner. The mixing chamber and burner should be carefully designed so the mixed gas is uniformly heated and distributed before it enters the catalyst bed. The catalyst bed is usually a deposit of platinum or another noble metal on the surface of a ceramic base that is shaped so that contact with the mixed gas is enhanced. Alternatively, the deposit may be on a metal mesh-pad structure. The catalyst bed depth is normally 12 in. but may be 8 to 24 in., and the volume may vary from 0.5 to 2 ft³ per 1000 scfm of flow through the bed.²⁻⁵

When heat recovery is practiced, the recuperative heat exchanger or waste-heat boiler is connected to the catalyst chamber. The hot flue gases leaving the catalyst are cooled by (1) the entering waste gas and/or combustion air being heated in the recuperative heat exchanger or (2) by the steam being produced in the waste-heat boiler.

If the waste gas is not under pressure in the process, a fan is usually required. If the waste gas does not contain at least 16 mole % oxygen, combustion air is required for the burner and another fan is required. When the VOC is mixed with air, the combustion air fan is not required.

The complete catalytic oxidizer system also has control instruments for temperature, flow, and fuel and for combustion safety. A stack is needed to exhaust the flue gases at sufficient height to be safely dispersed.⁴

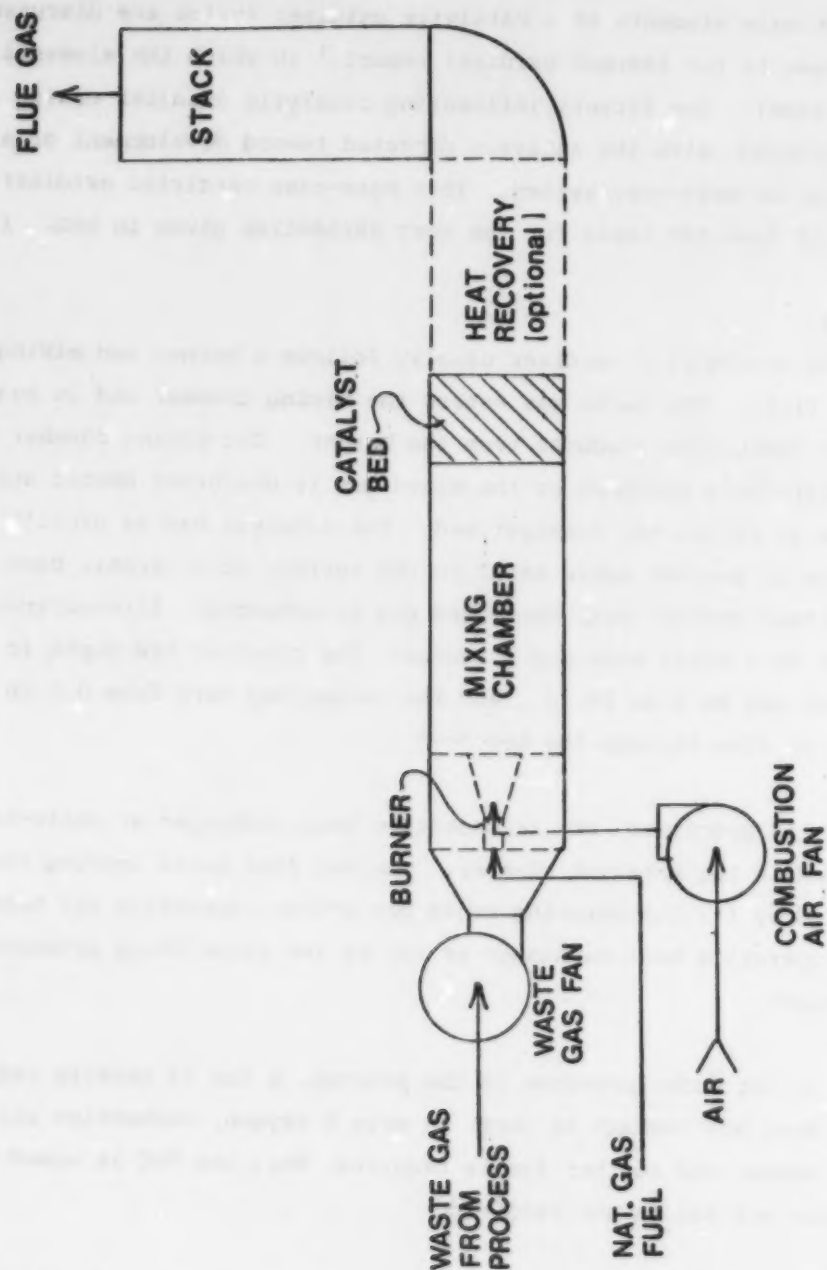


Fig. II-1. Basic Catalytic Oxidizer

The important variables in sizing a catalytic oxidizer system are the following:

1. the waste-gas flow rate
2. the waste-gas oxygen content
3. the VOC destruction efficiency
4. the waste-gas heat content, which is a function of the VOC concentration and composition
5. the operating temperatures

Except for item 5 the values for the parameters are the same as those used in the thermal oxidation report.¹ In addition to the case for which the waste-gas oxygen content is assumed to be zero, a study was made of the case in which the waste-gas oxygen content is the same as that for air.

The operating temperature conditions are more complex, with the following values assumed for this study:

1. the temperature of the waste gas from the process is 100°F,
2. the minimum temperature of the mixed gases entering the catalyst bed is 600°F to ensure an adequate initial reaction rate,
3. the minimum temperature of the flue gas leaving the catalyst bed is 900°F to ensure an adequate overall reaction rate to give the desired VOC destruction efficiency,
4. the maximum temperature of the flue gas leaving the catalyst bed is 1200°F (ref 5) to prevent catalyst deactivation by overheating,
5. the minimum temperature of the flue gases leaving the heat recovery section is 500°F to prevent condensation and corrosion of the heat-transfer surfaces.

For specific applications other temperatures may be appropriate. For example, temperatures of 400 to 450°F are reported at one installation and 500 to 550°F at another. The temperature of the flue gas leaving the catalyst bed is a function of VOC concentration and specific heat and may be less than 900°F for some waste gases that are more easily oxidized and still give the desired VOC destruction.⁶

Only 1 or 2% excess of oxygen may be required with an effective catalyst, although a minimum of 3 mole % oxygen in the flue gas is assumed for this study, as was used for the thermal oxidation study. By keeping air addition to the minimum required, supplemental fuel usage is minimized.⁶

Two levels of VOC destruction efficiency are assumed, 90% and 99%. See Table II-1 for the specific parameters used for the five waste-gas composition cases studied for this report.

1. Catalyst Bed Size

The amount of catalyst and the depth and cross-sectional area of the catalyst bed depend to a certain extent on the vendor's experience with waste gases having similar compositions to the one under consideration. When such experience is lacking, a pilot unit should be run on the actual stream for a period of time. For the base case of 90% destruction efficiency a catalyst volume of 1 ft³ per 1000 scfm of flue-gas flow is used; 1.5 ft³ of catalyst per 1000 scfm of flue-gas flow is assumed to give 99% destruction of VOC as indicated by Du Pont.⁵ They also indicate that the bed depth is normally 12 in. but can vary from 8 to 24 in.⁵ The degree of removal of combustible gaseous compounds depends on the mass-transfer limitations. A unit can be designed for a high degree of VOC removal by making it long enough (more catalyst), but this introduces a higher pressure drop through the device. In some cases therefore this increased pressure drop can be a limiting factor that dictates the degree of destruction.⁶

2. Mixing Chamber Design

The mixing chamber must be of sufficient length and design so that the flame from the preheat burner will not impinge on the catalyst bed and so that the waste gas will be well mixed and evenly heated before it enters the catalyst bed. The mixing chamber designs on which this study is based are those used by catalytic oxidizer vendors.^{5, 13-17} Details of the designs were not furnished. A typical design velocity of 25-35 fps for the heated mixed gases is usually adequate for mixing and is used for at least one type of catalyst bed.³

Table II-1. Parameters for Catalytic Oxidation Calculations^a

Waste-Gas Heat Content (Btu/scf)	No Heat Recovery or with Waste-Heat Boiler ^b				With Recuperative Heat Exchanger			
	Temperature (°F)		Ratio (Flue-Gas Flow Waste-Gas Flow)	Fuel Required (Btu/scf)	Temperature (°F)		Ratio (Flue-Gas Flow Waste-Gas Flow)	Fuel Required (Btu/scf)
	To Catalyst	From Catalyst			From Heat Exchanger	To Catalyst		
0	900	900	1.45	23.9	400	900	1.34	14.1
2	855	900	1.48	24.2	400	850	1.36	14.2
10	600	900	1.45	14.6	400	600	1.35	5.6
20	600	1150	1.58	15.9	600	600	1.39	0
10 (in air)	600	1019	1.0	10.1	600	600	1.0	0

^aCatalytic oxidizer conditions:

VOC molar heat of combustion = 730,250 Btu/lb-mole.

VOC molecular weight = 50.

VOC C, H, O fraction = 68.3 wt % C, 11.4 wt % H, 20.3 wt % O.

Waste gas temperature = 100°F.

Average waste-gas molecular weight = 29.

No oxygen in waste gas except where indicated "in air."

Water content of combustion air = 1.0 wt %.

3 mole % oxygen or greater in flue gas after oxidation.

^bTemperature of flue gas leaving waste-heat boiler = 500°F.

3. Fans

Fans for both the waste gas and the combustion air are required unless the waste gas contains sufficient oxygen (16 mole %) to be used in burning the auxiliary fuel. The pressure drops used in this report for catalytic oxidizer systems with conversions of 90% are 5 in. H₂O with no heat recovery and 14 in. H₂O with heat recovery, for either a recuperative heat exchanger or for a waste-heat boiler. For a system with 99% conversion, pressure drops of 7 in. H₂O with no heat recovery and 16 in. H₂O with heat recovery are used. The cost of the fans is included in the catalytic oxidizer costs given in Appendix A.

4. Recuperative Heat Exchangers

A recuperative heat exchanger transfers heat from the flue gas to the waste gas and combustion air, lowering the amount of fuel and combustion air required. See Sect. III-A-2 of the thermal oxidizer report¹ for a discussion of heat recovery by recuperative heat exchange. The heat content of the flue gas* from the catalytic oxidizers studied for this report is shown in Fig. II-2. The actual heat content of a flue gas depends on the relative quantities of fuel, air, VOC, and waste gas going to the catalytic oxidizer. The straight line of Fig. II-2 represents a best fit to several data points and is adequate for estimating the heat recovery and the surface area required. Figure II-3 shows the ratio of the heat exchanger surface to the flue-gas flow rate as a function of the percent of heat recovery and the overall heat-transfer coefficient (U) for recuperative heat exchangers for flue gases having temperatures of 900 to 1200°F. An overall heat-transfer coefficient of 4 Btu/(hr)(ft²)(°F) is used in this study.

5. Waste-Heat Boiler

For this study a waste-heat boiler producing 100-psig steam was chosen because the temperature of the flue gas from a catalytic oxidizer is 1200°F or less. A temperature of 500°F for the flue gas exhausted from the waste-heat boiler is used for this study. See Sect. III-A-4 in the thermal oxidizer report¹ for a discussion of heat recovery by use of a waste-heat boiler.

*For this report heat content of the flue gas refers to its sensible heat. Heat content of the waste gas refers to the energy generated by oxidation of the VOC contained in the waste gas.

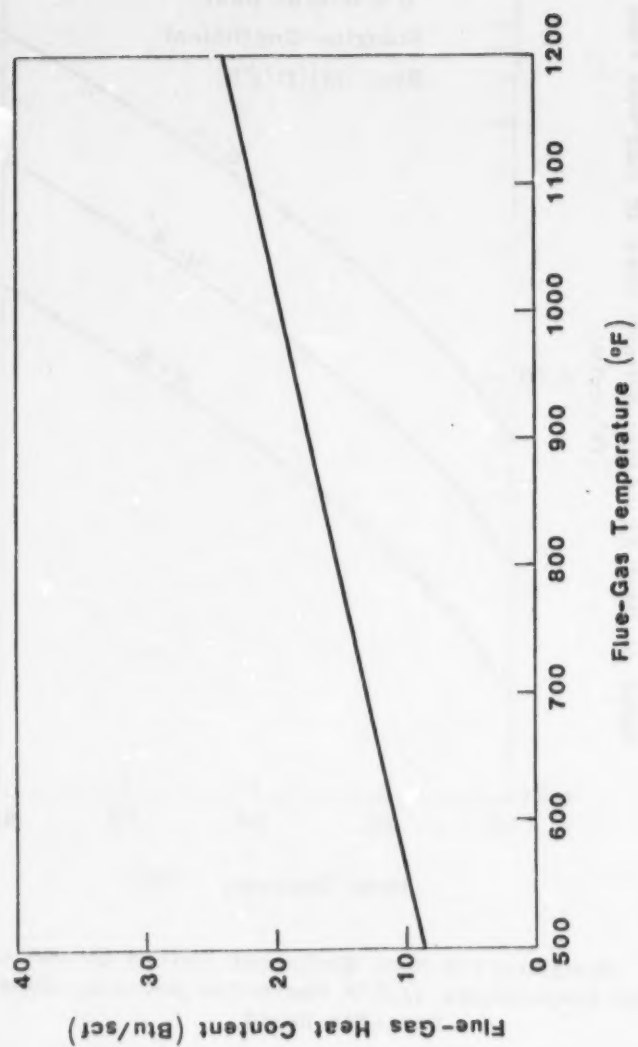


Fig. II-2. Flue-Gas Heat Content

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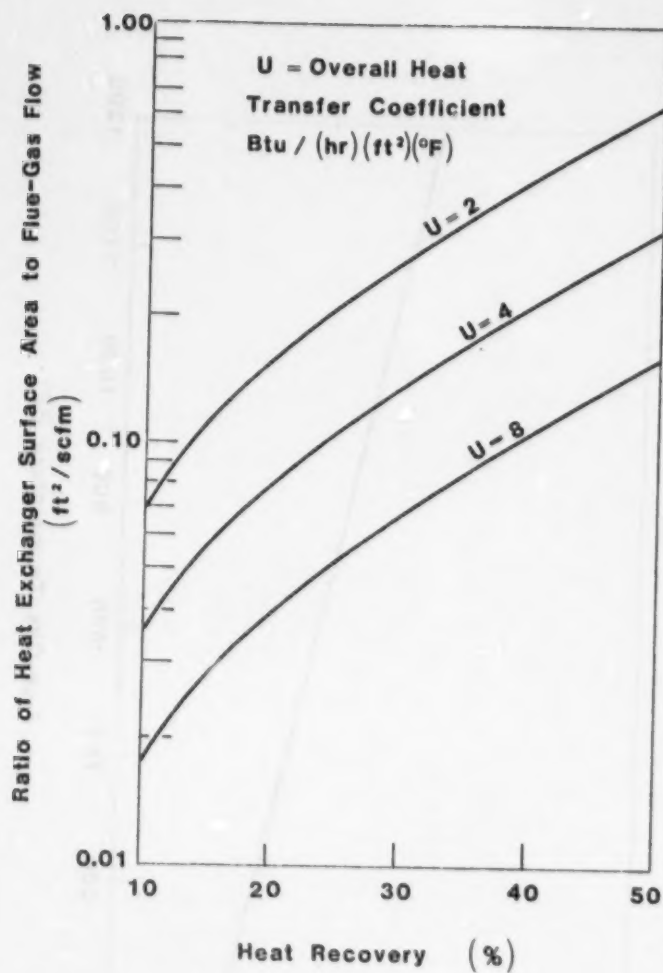


Fig. II-3. Recuperative Heat Exchanger Design at 900 to 1200°F
Flue-Gas Temperature (100°F Waste-Gas and Air, Constant
Specific Heat)

6. Stack and Other Equipment

The stack and ducting for this report are based on the same parameters used for the thermal oxidizer (see Sects. III-A-5 and 6 in the thermal oxidizer report¹).

B. CATALYTIC OXIDATION EFFICIENCIES

The destruction efficiency (conversion of VOC to carbon dioxide and water) for catalytic oxidizers depends on the catalyst type and volume, the operating temperature, and the composition and concentration of VOC in the waste gas. The chemical structure of a compound affects the destruction efficiency of a catalytic oxidizer more than it does a thermal oxidizer. A portion of the waste gas can by-pass the catalyst at times in most bed designs, and this limits the destruction efficiency that can be achieved.⁶ The catalyst may lose activity with time for such reasons as sintering, accumulations of poisons, or an accumulation of products. Although a higher operating temperature may compensate to some extent for this loss in activity, there is a limit dictated by the maximum operating temperature of the catalyst. In some cases the catalyst may have to be cleaned periodically for activity and destruction efficiency to be maintained. The actual life of the catalyst will vary with application and is uncertain for a new use. Some catalysts are reported to have been in use for over 8 years, but in some applications the catalyst must be replaced every year; the average life appears to be in the range of 3 to 5 years. For this study replacement of the catalyst every three years is assumed.^{2—5}

III. CONSIDERATIONS FOR INSTALLATION OF CATALYTIC OXIDIZERS

A. NEW PLANTS

All requirements that are considered when thermal oxidizers are installed also apply to the installation of catalytic oxidizers (see Sect. IV of the thermal oxidizer report¹).

B. EXISTING PLANTS

All considerations for installation of a catalytic oxidizer in a new plant also apply to retrofitting one in an existing plant. However, if the steam generation boiler in an existing plant is adequate, a waste-heat boiler may not be economically feasible. The costs and cost-effectiveness data presented in this report are not intended to apply to retrofitted catalytic oxidizer systems. In retrofitted systems additional costs may be encountered because of such items as demolition requirements, crowded construction working conditions, scheduling construction activities with production activities, and longer interconnecting ducts. These factors are site-specific, and no attempt has been made to provide costs. For specific retrofit cases, rough costs may be obtained by using the new-site data and adding as required for a specific retrofit situation.

IV. COST AND ENERGY IMPACTS OF CATALYTIC OXIDATION

A. BASE-CASE CATALYTIC OXIDIZER DESIGN SUMMARY

The results of cost- and energy-effectiveness calculations for the typical or base-case catalytic oxidation system are presented here. The catalytic oxidizer conditions, the waste-gas heat contents, and the resulting temperatures given in Table II-1 for the base-case design establish the fuel requirement, the ratio of flue-gas flow to waste-gas flow, and the percent of heat recovered. Costs are estimated for seven waste-gas flows: 700, 2,000, 5,000, 10,000, 20,000, 50,000, and 100,000 scfm; for destruction efficiencies of 90 and 99%; and for no heat recovery, heat recovery with a recuperative heat exchanger used to heat the waste gas and combustion air, and heat recovery with a waste-heat boiler used to produce steam.

B. COST BASIS

The estimated capital costs for the catalytic oxidation systems described represent the total investment required for purchase and installation of all equipment and material to provide a facility like that described in Sect. II. This includes all indirect costs, such as engineering and contractors' fees and overheads. The estimated capital costs are battery-limits costs and do not include provision for bringing utilities, services, or roads to the site, backup facilities, land, research and development, or process piping and instrumentation interconnections that may be required within the process generating the waste-gas feed to the catalytic oxidation system.

The method used to develop the estimated capital costs was based on applying certain factors to the purchase prices of equipment to arrive at an installed capital cost. Purchase costs were obtained from vendors and previous EPA reports as described below. Table IV-1 gives the ranges used for factoring up the purchased price of equipment to the installed cost and is based on historical data of IT Enviroscience Process Engineering. The expected accuracy of the total installed cost thus obtained is in the range of $\pm 30\%$. This method of obtaining total installed capital costs is suitable for study or screening estimates.

For catalytic oxidation systems a 30% allowance was added to the estimated major equipment purchase cost to compensate for unspecified items, resulting in the

Table IV-1. Factors Used for Estimating Total Installed Costs

A = Major Equipment Purchase Cost Plus 0.1 to 0.35 Allowance

Installation costs

Foundations	0.06A + \$100 X number of pumps
Structures	0.15A (no structures) to 0.30A (multideck structures)
Equipment Erection	0.15A to 0.30A (depending on complexity)
Piping	0.40A (package units) to 1.10A (rat's nest)
Insulation	0.06A or 0.15 X piping (normal) to 0.30 X piping (bulk hot or cold)
Paint	0.05A
Fire Protection	0.01A to 0.06A (depending on requirements)
Instruments	0.10A to 0.30A or 0.01A to 0.25A + \$50,000 to \$300,000 for process control computer
Electrical	0.15A or 0.05A + \$500 per motor
B = Base Cost	A + Sum of Installation Costs
Sales Tax	0.025A + 0.025B
Freight	0.16A
Contractor's Fees	0.30 (B-A)
C = Total Contract	B + Taxes, Freight, and Fees
Engineering ^a	0.01C to 0.20C
Contingencies ^b	0.15C
D = Unit Installed Cost	C + Engineering + Contingencies
E = Total Subestimates	Sum of semidetalled subestimates (buildings, site development, cooling towers, etc.). Each subestimate should include taxes, freight, fees, engineering and contingency, and should be escalated to date of expenditure for that cost component. Engineering costs, contingencies, and escalation factors for these subestimates will vary according to the type of job.
F = Total Project Cost	D + E

^aIncludes cost from capital project teams, process engineering, engineering, purchasing, and other support groups.

^bContingency should not be applied to any cost component that has been committed by either purchase order or contract.

total estimated equipment purchase cost designated A in Table IV-1. This established the basis for the application of all the installed capital cost factors shown.

The sum of the installation costs for a catalytic oxidizer without a catalyst bed or a connecting duct and with a minimum stack was estimated to be about 1.1 times A. The base cost (B) is therefore approximately 2.1 times the total estimated equipment purchase cost (A). Additional percentages were applied to the base cost (B) as shown in Table IV-1 to arrive at a unit installed cost (D) for a catalytic oxidizer with a minimum stack but without a catalyst bed and without a connecting duct to the process. The initial catalyst installed cost (D) was estimated by multiplying the catalyst purchase cost obtained from Fig. A-2 by 1.2 to account for the costs of freight, taxes, and labor and fees for installing the catalyst bed in the oxidizer.

The sum of the installation costs for a waste heat boiler was also estimated to be about 1.1 times A. For a recuperative heat exchanger, for the connecting duct from the process to the catalytic oxidizer, and for the 80-ft stack the sum of the installation costs was estimated to be about 0.5 times A (B is therefore approximately 1.5 times A) because the foundations, structures, erection, and piping will be incremental and relatively small when compared to those required for the catalytic oxidizer. The same additional percentages for sales tax, freight, contractor's fees, engineering, and contingencies were applied to the base cost (B) to arrive at the unit installed costs (D) (see Table IV-1). Allowances for the cost of site development and for the cost of vendor assistance during startup were added to the total of all the unit installed costs (D) for each case to give the total installed capital costs shown in Figs. IV-1 through IV-3.

The basic pieces of equipment are catalytic oxidizers, recuperative heat exchangers, and waste-heat boilers, plus the catalyst, ducts, and stacks. The sources of purchase cost data for these items follow, and curves showing purchase costs of the catalytic oxidizers, the catalyst, and the waste-heat boilers are given in Appendix A. A curve showing the purchase costs of recuperative heat exchangers is shown in Appendix A of the thermal oxidizer report.¹

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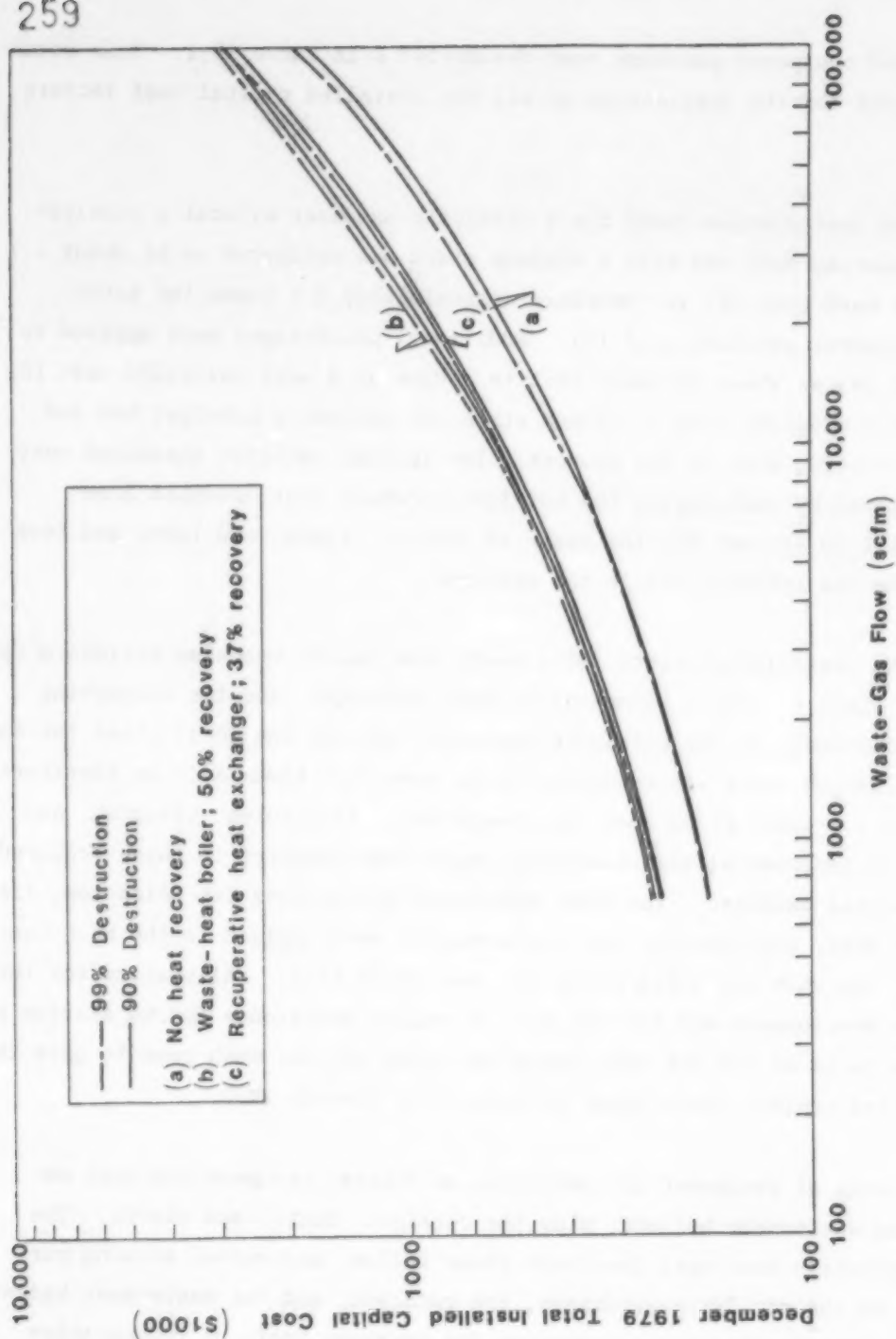


Fig. IV-1. Installed Capital Costs of Catalytic Oxidizer Systems for Waste Gas with Heat Contents of 0 to 10 Btu/scf

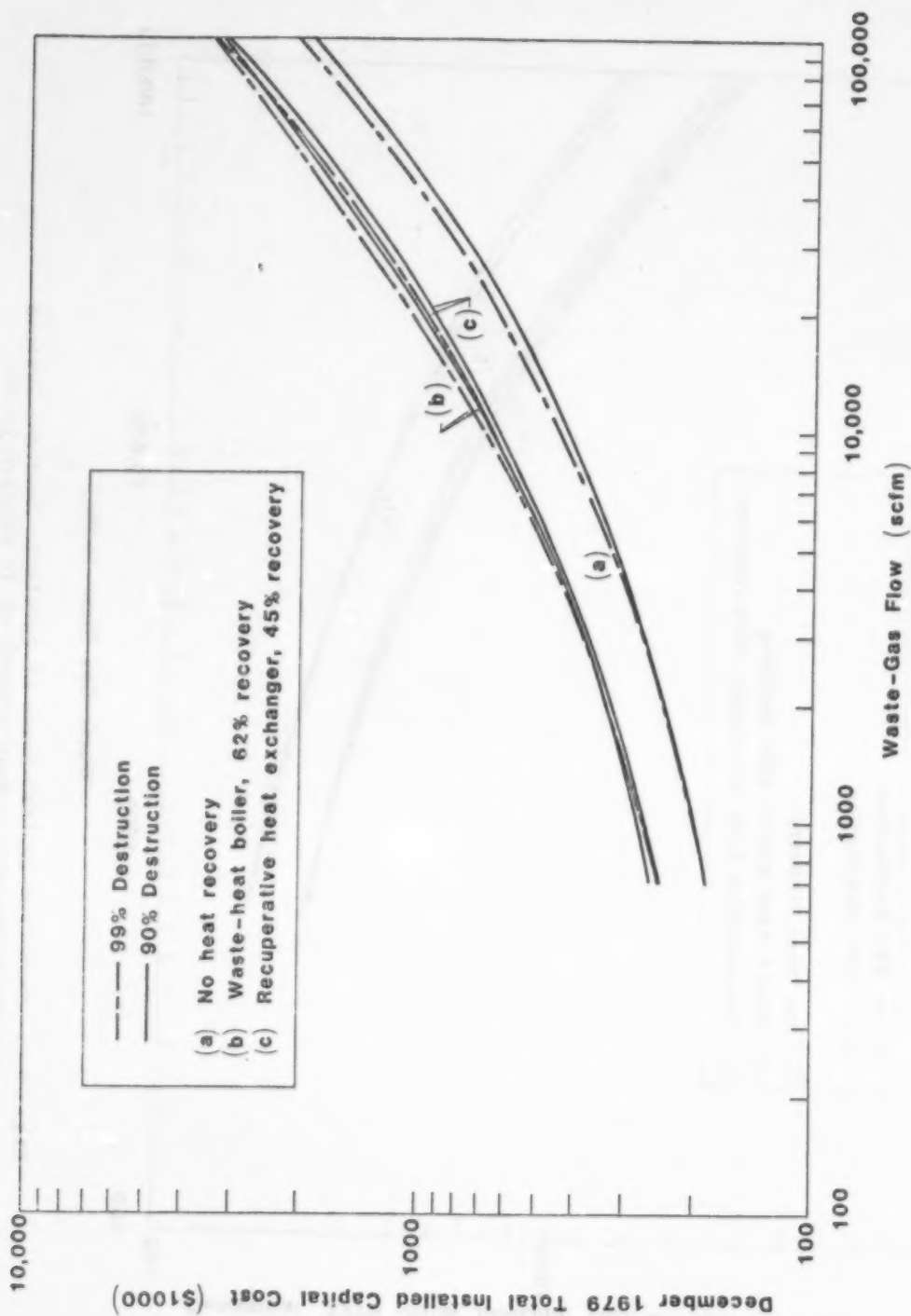


Fig. IV-2. Installed Capital Costs of Catalytic Oxidizer Systems for Waste Gas with a Heat Content of 20 Btu/scf

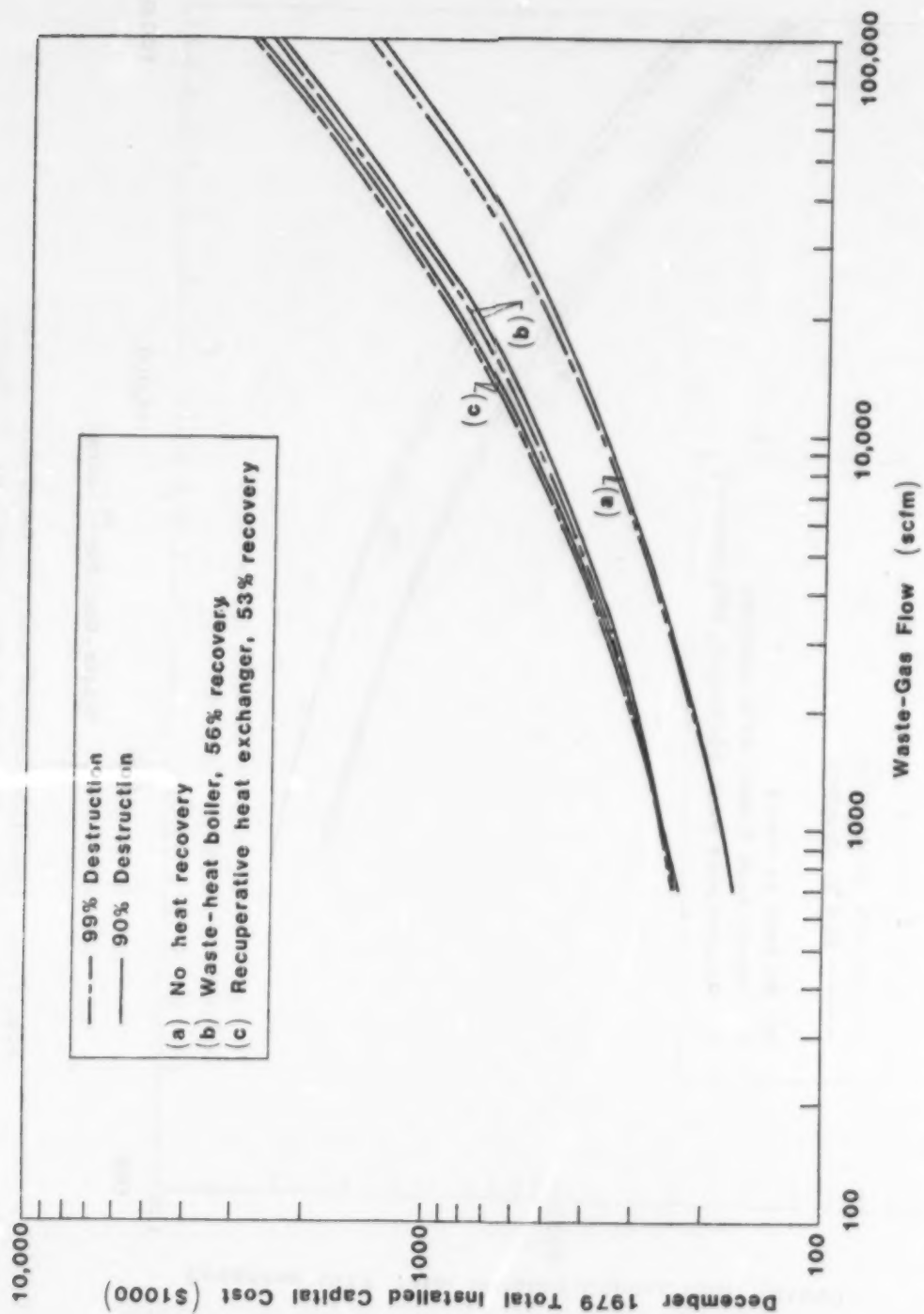


Fig. IV-3. Installed Capital Costs of Catalytic Oxidizer Systems for Waste Gas with a Heat Content of 10 Btu/scf in Air

1. Purchase Costs of Catalytic Oxidizers

Preliminary purchase costs for catalytic oxidizers were obtained from vendors. Several vendors were contacted by telephone and letter and three supplied cost data and other information: Du Pont,⁵ Englehard,¹³ and Oxy-Catalyst.¹⁴ The costs were for prepiped and prewired units complete with the burner, blower, refractory, controls, etc., required for handling various waste-gas flows [in standard cubic feet per minute (scfm)]. Purchase cost data were also extracted from previous EPA reports and escalated to December 1979. The curves shown in Appendix A for the purchase costs of the catalytic oxidizers and of the catalyst were derived from the combined data.

2. Purchase Costs of Recuperative Heat Exchangers

The purchase costs of recuperative heat exchangers were obtained from Appendix A of the thermal oxidizer report.¹

3. Purchase Costs of Waste-Heat Boilers

The purchase costs of waste-heat boilers for 100-psig steam were estimated by adjusting the purchase costs from the thermal oxidizer report¹ to compensate for the differences in flue-gas temperatures and steam pressures.

4. Purchase Costs of Ducts

Each system is assumed to require 150 ft of round-steel inlet ductwork with the same fittings shown for the thermal oxidizer.¹ The costs used are also the same.

5. Purchase Costs of Stacks

The costs of the stacks are the same as those used for thermal oxidizers.¹

C. ANNUAL COSTS

Annual costs for various operating conditions are given in Appendix B. These costs are the basis for all the net annual cost graphs included in the report. The basis used in calculating these annual costs is defined in Table IV-2.

Figures IV-4 through IV-9 present the net annual costs of catalytic oxidation for various cases.

Table IV-2. Annual Cost Parameters

Operating factor	8760 hr/yr ^a
Operating labor	\$15/man-hr
Fixed Costs	
Maintenance labor plus materials, 6%	} 29% of installed capital
Capital recovery, 18% ^b	
Taxes, insurances, administration charges, 5%	
Utilities	
Electric power	\$0.03/kWh
Natural gas	\$2.00/million Btu
Heat recovery credits (equivalent to natural gas)	\$2.00/million Btu

^aControl devices will usually operate on the same cycle as the process. Process downtime is normally expected to range from 5 to 15%. If the hourly rate remains constant, the annual production and annual VOC emissions will be correspondingly reduced. From the standpoint of cost-effectiveness calculations, the error introduced by assuming continuous operation is negligible.

^bBased on 10-year life and 12% interest.

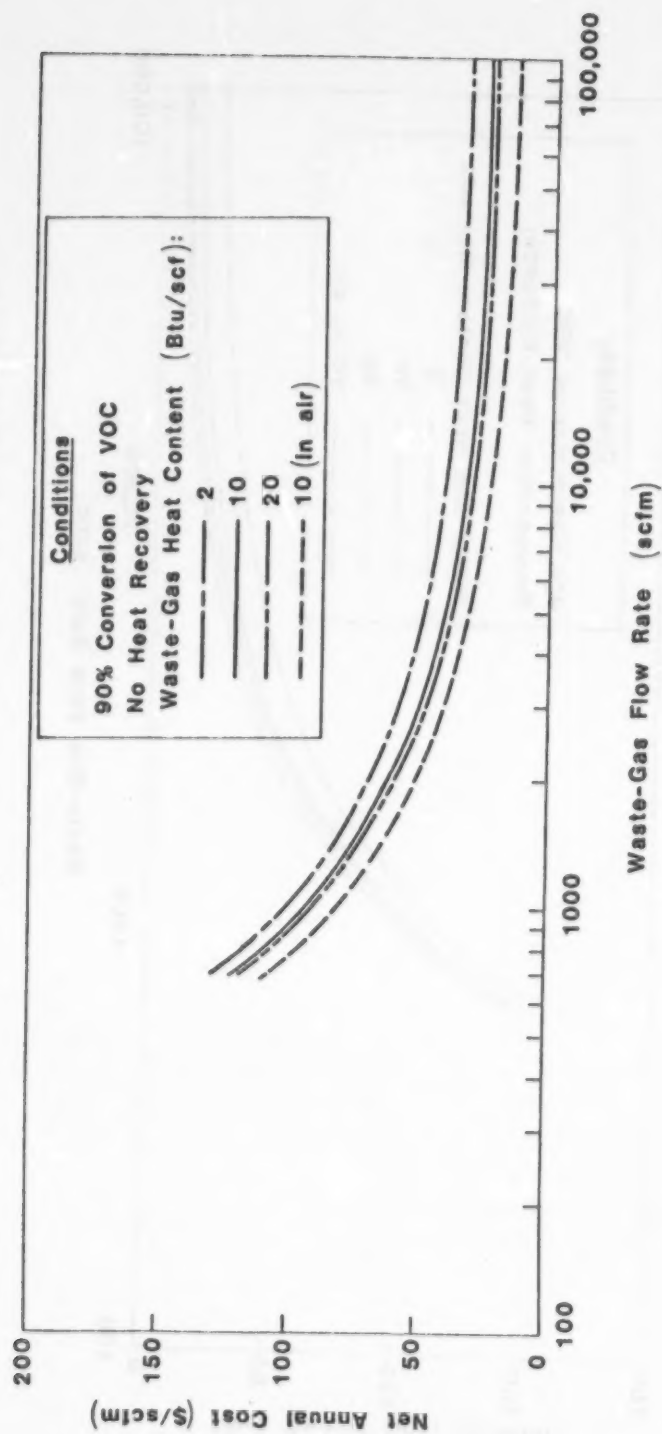


Fig. IV-4. Net Annual Costs vs Waste-Gas Flow Rate for Catalytic Oxidizers Having a Destruction Efficiency of 90% and with No Heat Recovery

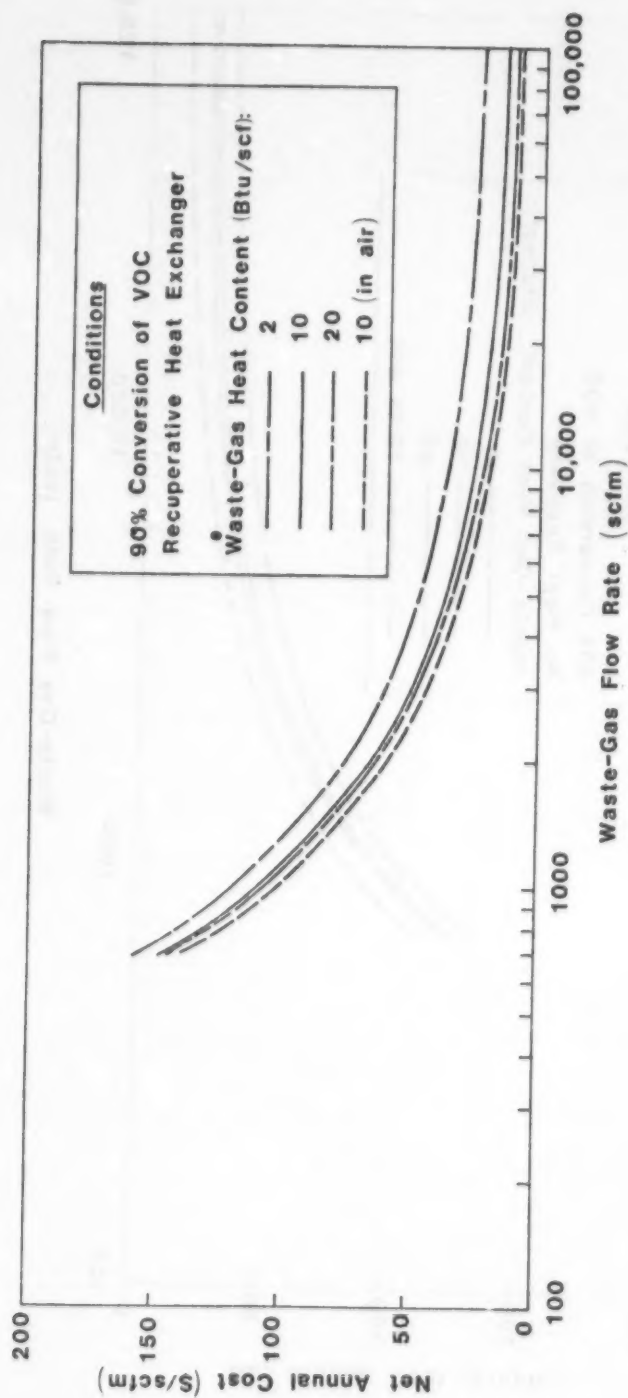


Fig. IV-5. Net Annual Costs vs Waste-Gas Flow Rate for Catalytic Oxidizers Having a Destruction Efficiency of 90% and with Recuperative Heat Exchangers

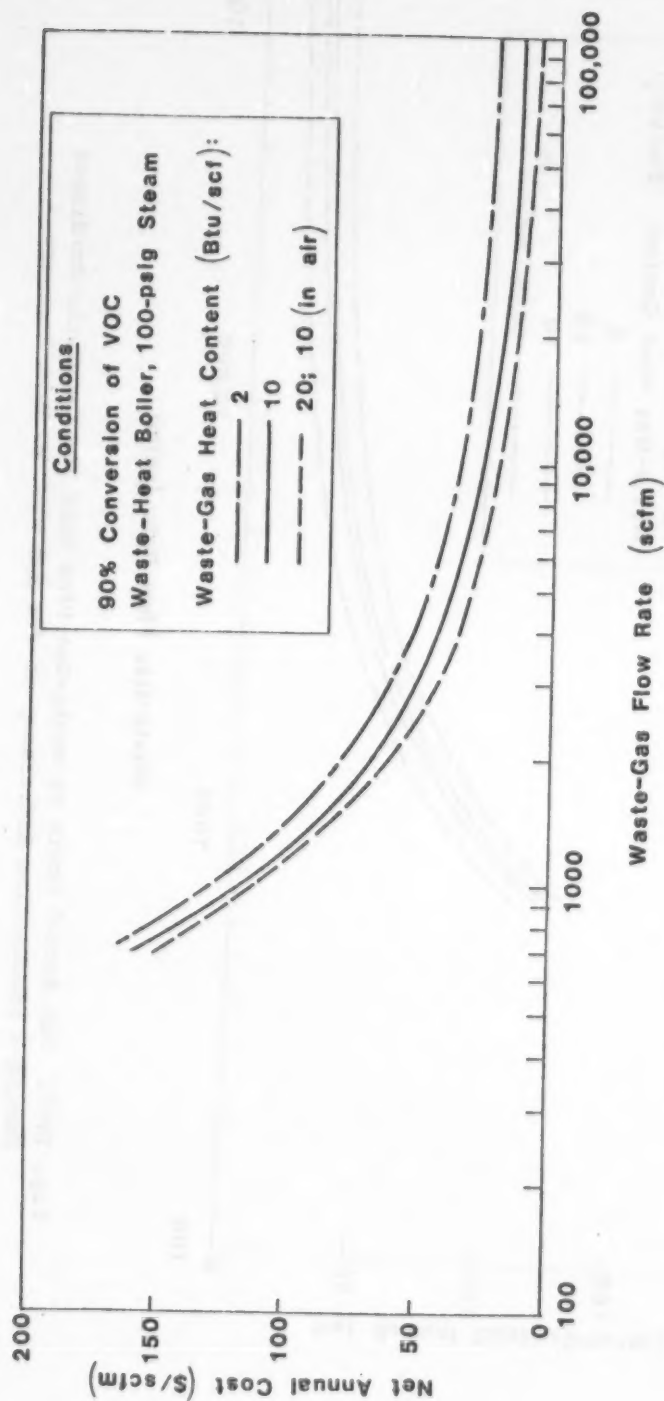


Fig. IV-6. Net Annual Costs vs Waste-Gas Flow Rate for Catalytic Oxidizers Having a Destruction Efficiency of 90% and with Waste-Heat Boilers

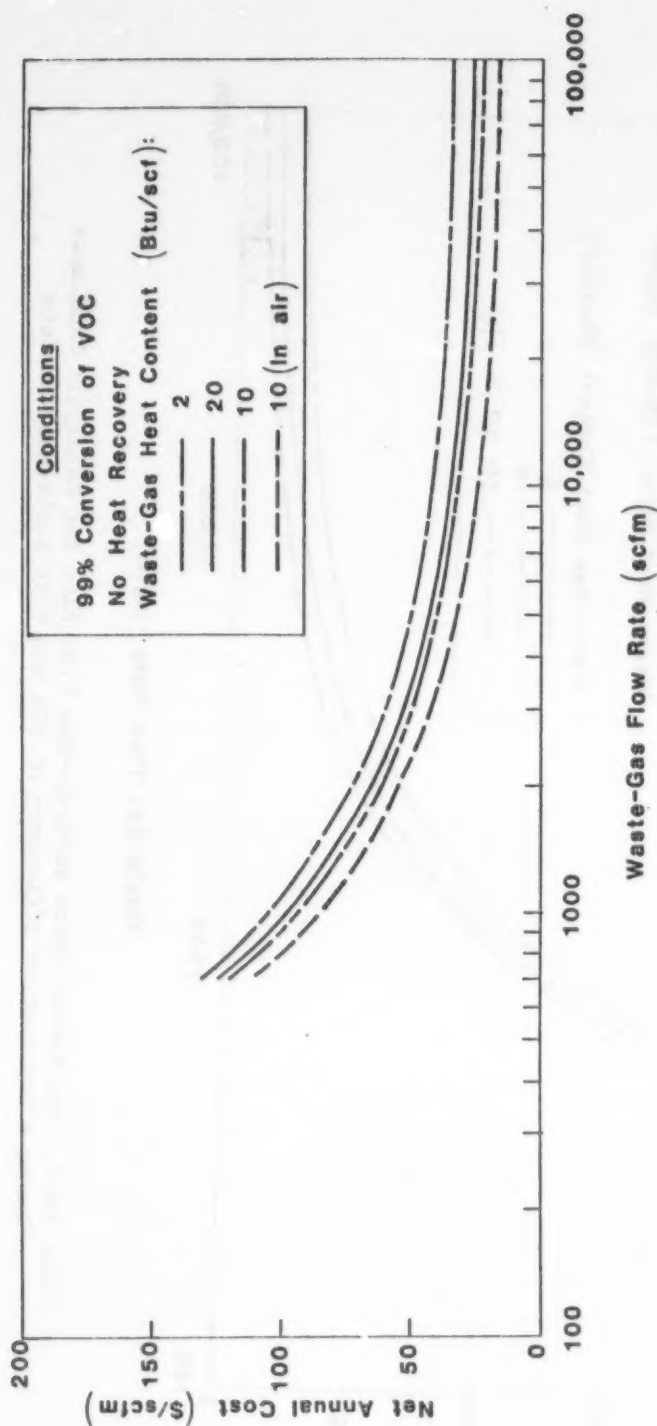


Fig. IV-7. Net Annual Costs vs Waste-Gas Flow Rate for Catalytic Oxidizers Having a Destruction Efficiency of 99% and with No Heat Recovery

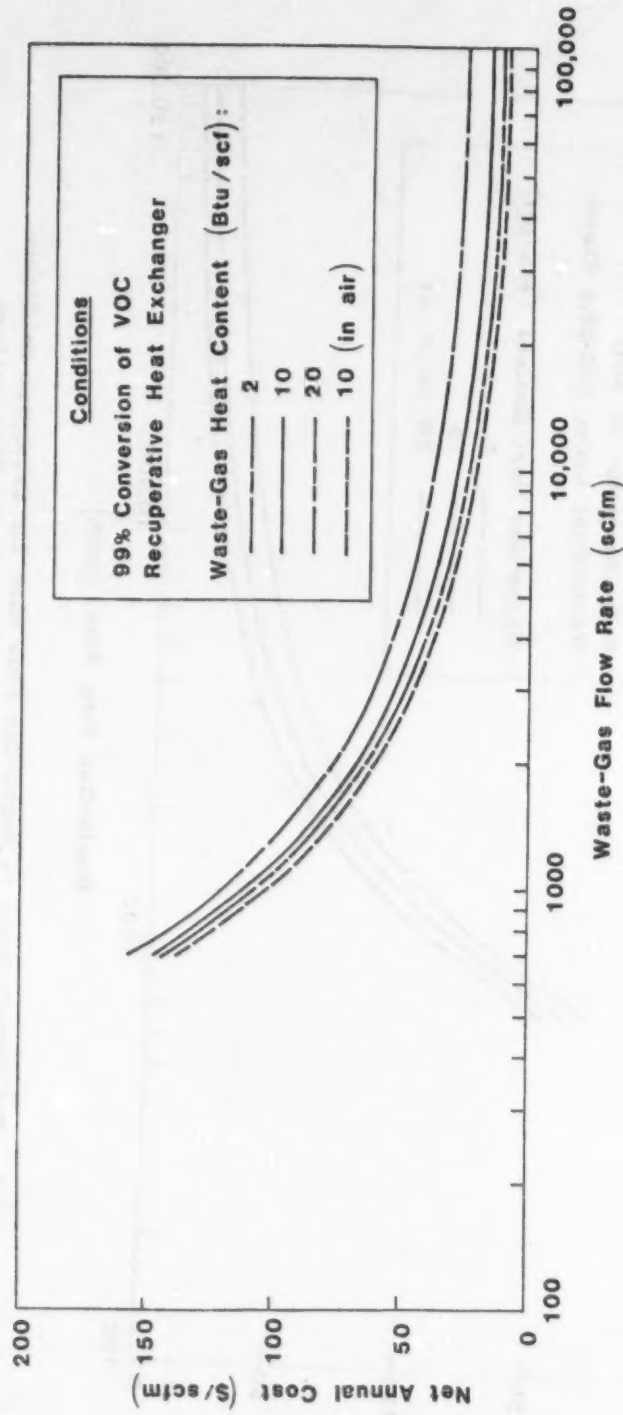


Fig. IV-8. Net Annual Costs vs Waste-Gas Flow Rate for Catalytic Oxidizers Having a Destruction Efficiency of 99% and with Recuperative Heat Exchangers

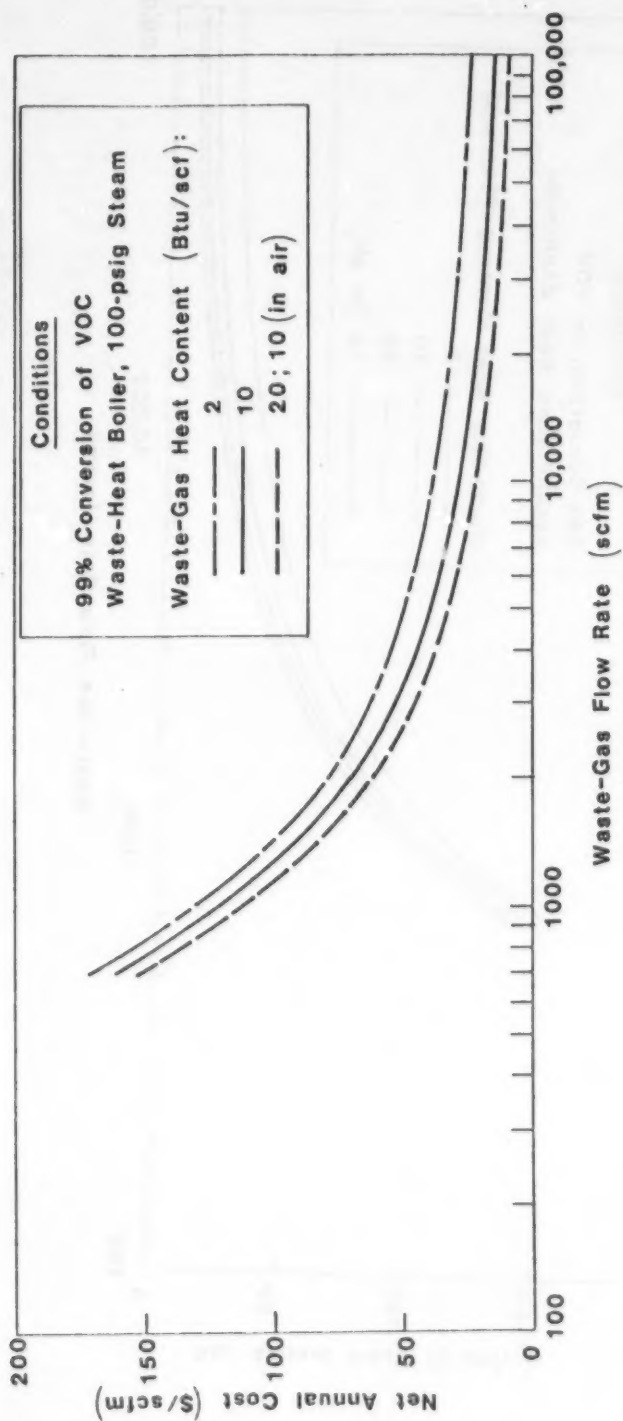


Fig. IV-9. Net Annual Costs vs Waste-Gas Flow Rate for Catalytic Oxidizers Having a Destruction Efficiency of 99% and with Waste-Heat Boilers

D. COST AND ENERGY EFFECTIVENESS

The cost effectiveness and energy effectiveness were calculated by dividing the annual cost for a particular option (Appendix B) or the fuel usage in Btu/yr by the total annual amount of VOC destroyed with the conversion efficiencies assumed in Sect. II. The cost effectiveness is given in Table IV-3 and the energy effectiveness is given in Table IV-4. Data on cases not shown in the cited tables can be easily developed by use of Appendix B.

Table IV-3. Cost Effectiveness of Catalytic Oxidation

Waste-Gas Heat Content (Btu/scf)	Waste-Gas Flow Rate (scfm)	VOC Destroyed (lb/hr) at 90% ^a	VOC Destroyed (lb/hr) at 99% ^a	Cost Effectiveness (per lb of VOC Destroyed) ^a					
				90% VOC Destruction ^b			99% VOC Destruction ^c		
				No Heat Recovery	Recuperative Heat Exchanger	Waste-Heat Boiler	No Heat Recovery	Recuperative Heat Exchanger	Waste-Heat Boiler
2	700	5.18	5.69	\$2.00	\$2.37	\$2.63	\$1.84	\$2.18	\$2.41
	5,000	37.0	40.7	0.758	0.733	0.735	0.710	0.685	0.688
	50,000	370	407	0.517	0.410	0.389	0.489	0.391	0.373
	100,000	739	813	0.503	0.390	0.362	0.476	0.372	0.348
10	700	25.9	28.5	0.367	0.445	0.495	0.338	0.409	0.454
	5,000	185	203	0.120	0.118	0.116	0.113	0.111	0.110
	50,000	1,850	2,030	0.0719	0.0538	0.0472	0.0691	0.0524	0.0467
	100,000	3,700	4,070	0.0692	0.0497	0.0417	0.0665	0.0486	0.0416
20	700	51.8	56.9	0.188	0.219	0.239	0.173	0.201	0.219
	5,000	370	407	0.0632	0.0529	0.0481	0.0596	0.0500	0.0459
	50,000	3,700	4,070	0.0388	0.0200	0.0127	0.0373	0.0203	0.0135
	100,000	7,390	8,130	0.0375	0.0180	0.00925	0.0361	0.0181	0.0104
10 (in air)	700	25.9	28.5	0.336	0.423	0.466	0.309	0.387	0.427
	5,000	185	203	0.0977	0.0985	0.0961	0.0916	0.0924	0.0901
	50,000	1,850	2,030	0.0518	0.0348	0.0297	0.0497	0.0342	0.0296
	100,000	3,700	4,070	0.0488	0.0305	0.0242	0.0469	0.0302	0.0246

^aVOC molecular weight = 50; molar heat of combustion = 730,250 Btu/lb-mole.^b1 ft³ of catalyst per 1000 scfm.^c1.5 ft³ of catalyst per 1000 scfm.

Table IV-4. Fuel Energy Effectiveness of Catalytic Oxidation

Waste-Gas Heat Content (Btu/scf)	Fuel Energy Usage (Btu/scf)			Net Energy Usage with Waste-Heat Boiler (Btu/scf)			Energy Effectiveness (Btu/lb of VOC Destroyed) ^a					
	Fuel Energy Usage (Btu/scf)			Net Energy Usage with Waste-Heat Boiler (Btu/scf)			90% VOC Destruction ^b			99% VOC Destruction ^c		
	No Heat Recovery of Waste-Heat Boiler	Recuperative Heat Exchanger	Waste-Heat Boiler	No Heat Recovery of Waste-Heat Boiler	Recuperative Heat Exchanger	Waste-Heat Boiler	No Heat Recovery	Recuperative Heat Exchanger	Waste-Heat Boiler	No Heat Recovery	Recuperative Heat Exchanger	Waste-Heat Boiler
2	24.2	14.2		11.8			196,000	115,000	96,000	179,000	105,000	87,000
10	14.6	5.6		2.2			23,700	9,090	3,570	21,500	8,260	3,250
20	15.9	0		(6.1)			12,900	0	(4,900)	11,700	0	(4,400)
10 (in air)	10.1	0		(1.0)			16,400	0	(1,620)	14,900	0	(1,480)

^aVOC molecular weight = 50; molar heat of combustion = 730,250 Btu/lb-mole.^b1 ft³ of catalyst per 1000 scfm.^c1.5 ft³ of catalyst per 1000 scfm.

V. SUMMARY AND CONCLUSIONS

Catalytic oxidation is used in several industries as a control technique for VOC emissions. The VOC in the off-gases from several processes in the synthetic organic chemical manufacturing industry are controlled by use of catalytic oxidation. A design criterion and design procedures are presented that allow for a preliminary catalytic oxidation design. A catalytic oxidation system with two destruction efficiencies and three heat recovery options, i.e., no heat recovery, heat recovery with a recuperative heat exchanger, or heat recovery with a waste-heat boiler, is considered. Capital and operating costs are developed, and the annual cost of catalytic oxidation is calculated as a function of the characteristics of the waste gas. The cost effectiveness and energy effectiveness of the two VOC destruction efficiencies and of three heat recovery cases are developed.

The conclusions of the cost evaluation are as follows:

1. The waste-gas flow rate is a highly sensitive variable in the determination of the annual cost and cost effectiveness (see Table IV-3). Energy effectiveness is independent of flow rate. As the feed flow rate increases, the annual costs increase but the annual cost per scfm decreases. The annual cost per scfm decreases quickly between 700 and 4000 scfm. The ratio decreases moderately between 4,000 and 40,000 scfm and is almost constant above 40,000 scfm.
2. The cost effectiveness and the energy effectiveness are strongly dependent on the waste-gas heat content, i.e., the VOC concentration (see Table IV-3, Table IV-4, and Appendix B). At 20 Btu/scf or 10 Btu/scf in air no fuel is required when recuperative heat recovery is used and the system is at operating temperature so that the energy effectiveness is almost zero. At the same VOC concentrations the recovery credit is greater than the fuel energy required if a waste-heat boiler is employed.
3. The annual cost, cost effectiveness, and energy effectiveness are slightly sensitive to the destruction efficiency or amount of catalyst (see Tables IV-3 and IV-4 and Figs. IV-4 through IV-9). This sensitivity is greater at low waste-gas flow rates than at high flow rates.

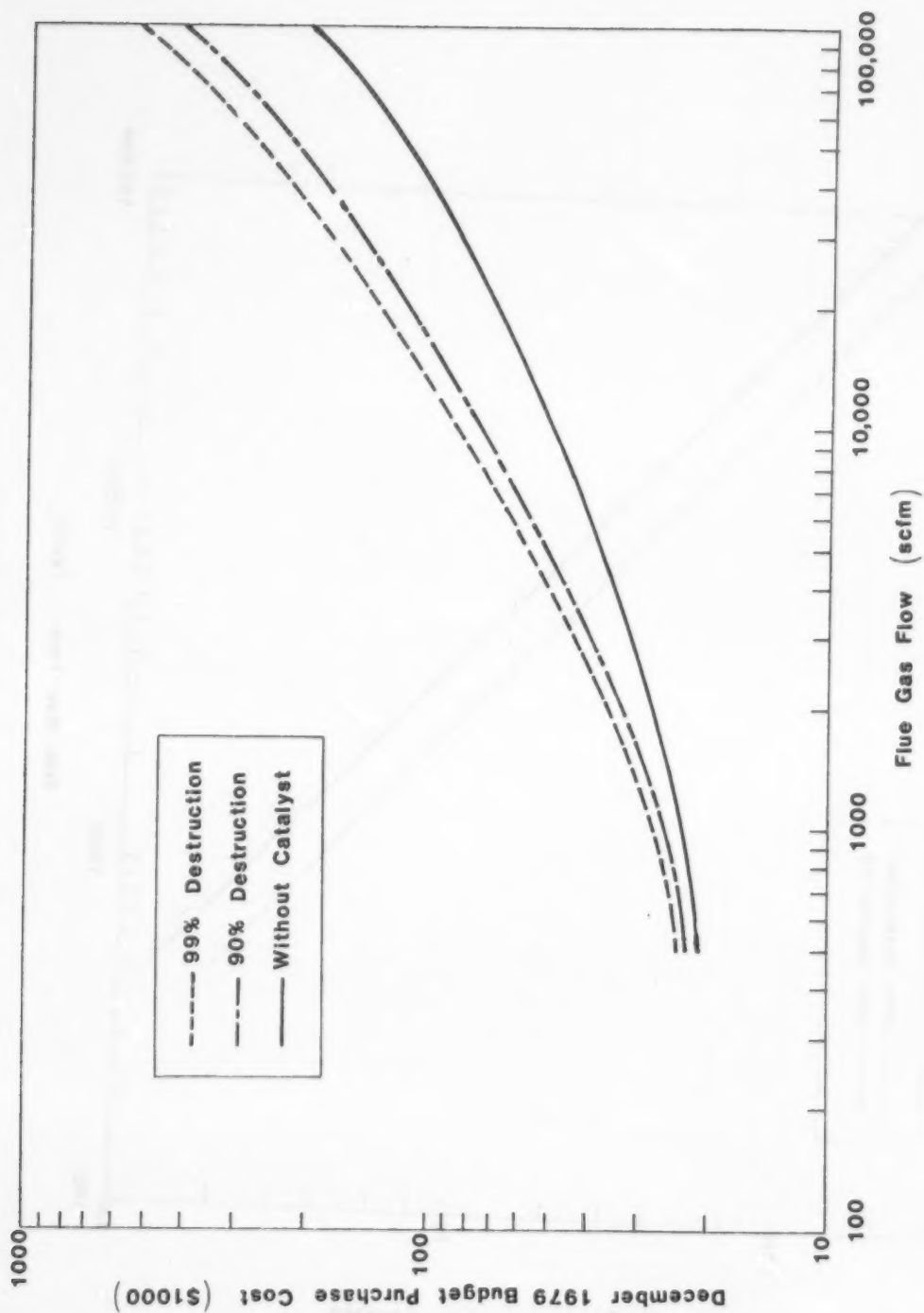
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15. CE Air Preheater, Industrial Gas Cleaning Institute, Report of Fuel Requirements, Capital Cost and Operating Expense for Catalytic and Thermal Afterburners, EPA-450/3-76-031, Research Triangle Park, NC (September 1976).

16. M. L. Kinkley and R. B. Neveril, Gard, Inc., Capital and Operating Costs of Selected Air Pollution Control Systems, EPA-450/3-76-014, Research Triangle Park, NC (May 1976).
17. Industrial Gas Cleaning Institute, Air Pollution Control Technology and Costs in Seven Selected Areas, EPA 450/3-73-010, Research Triangle Park, NC (December 1973).

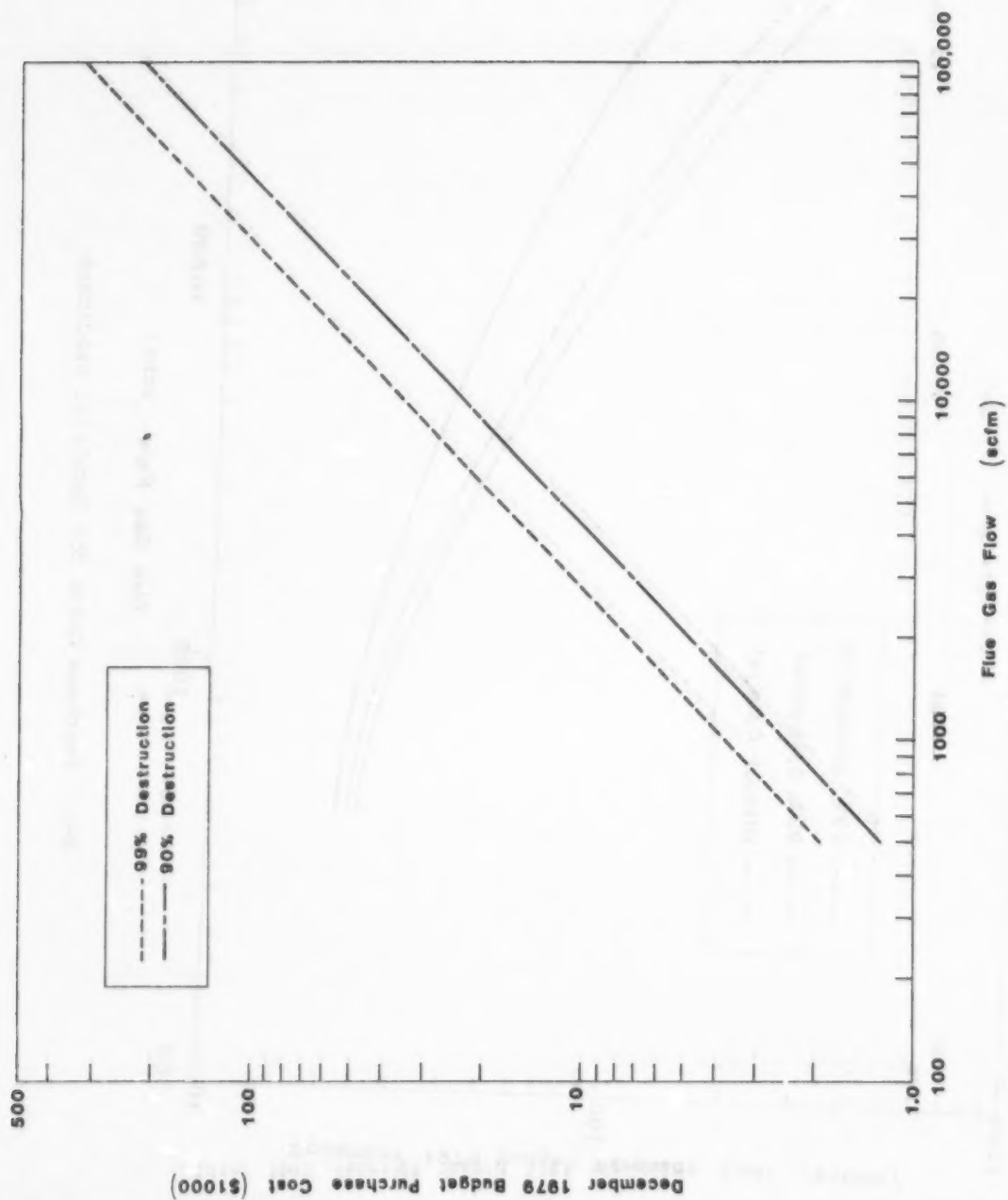
*Usually, when a reference is located at the end of a paragraph, it refers to the entire paragraph. If another reference relates to certain portions of that paragraph, that reference number is indicated on the material involved. When the reference appears on a heading, it refers to all the text covered by that heading.

APPENDIX A
PURCHASE COSTS FOR CATALYTIC OXIDIZERS,
CATALYST, AND WASTE-HEAT BOILERS



A-1. Purchase Costs for Catalytic Oxidizers

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A-2. Purchase Costs for Catalysts

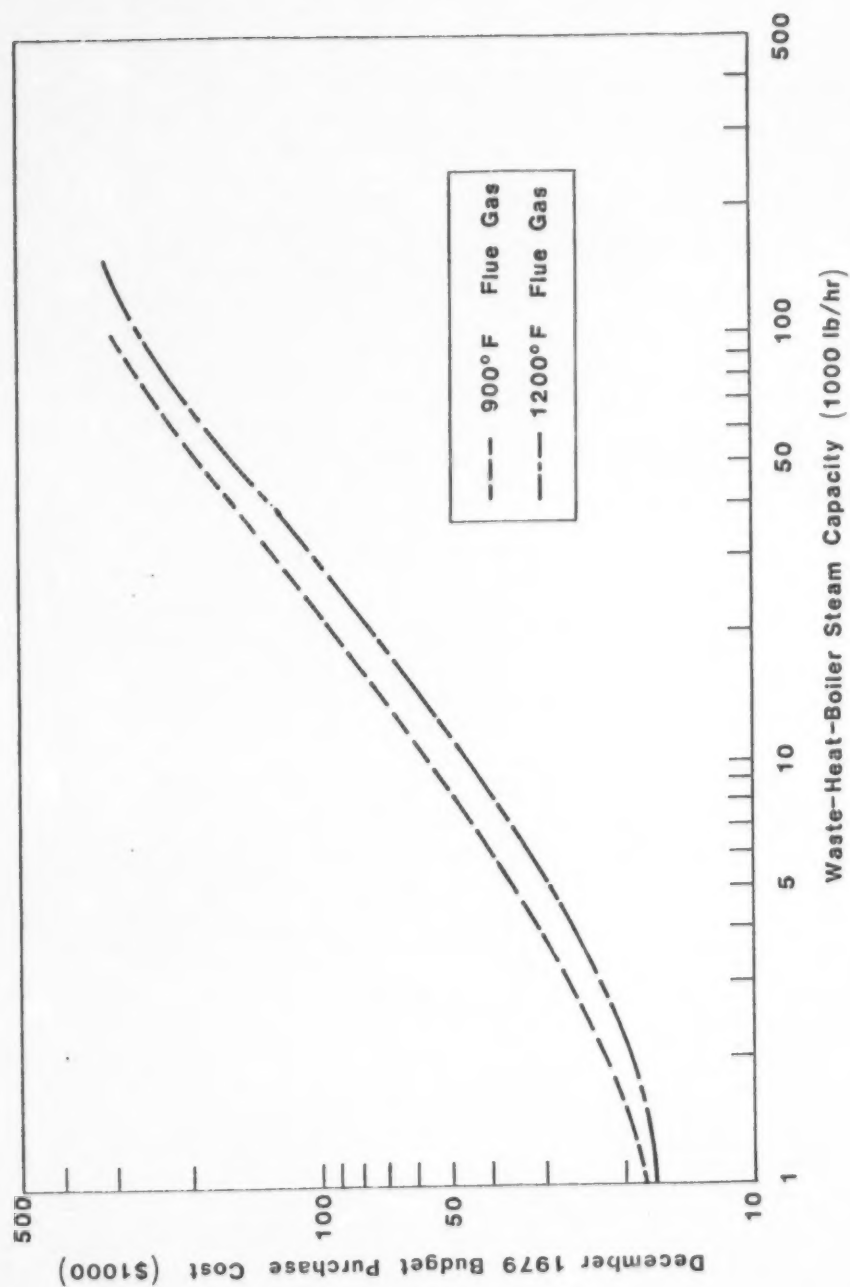


Fig. A-3. Purchase Cost for Waste-Heat Boilers Producing 100-psig Steam

APPENDIX B
ANNUAL COST DATA

SAMPLE CALCULATIONS—ANNUAL COST DATA

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The following sample calculations are based on an off-gas stream having a heat content of 10 Btu/scf and consisting primarily of nitrogen (combustion air must be supplied to the preheat burner and air must also be mixed with the off-gas so that after combustion of the fuel to the burner and oxidation of the VOC in the off-gas the flue gas contains 3 mole % oxygen).

Basis of calculations:

Off-gas flow rate	8,000 scfm
Off-gas temperature	100°F
Destruction of VOC	99%
Flue gas to off-gas ratio	1.45 scf/scf (Table II-1)
Heat recovery	Waste-heat boiler generating 100-psig steam with 50% heat recovery and flue gas in at 900°F and out at 500°F

Capital cost = \$563,000; from Fig. IV-1.

Fixed cost = (\$563,000 X 0.29*) + catalyst replacement cost (\$16,000) = \$179,000/yr.

Catalyst replacement cost =

$$\frac{\text{catalyst purchase cost } (\$40,000) \times \text{installation factor } (1.2)}{3\text{-yr replacement}}$$

Catalyst purchase cost = \$40,000; from Fig. A-2
(1.45 X 8000 scfm = 11,600 scfm of flue gas).

Operating cost = fuel (\$122,780) + electricity (\$9,551) + labor (\$36,000)
= \$168,331/hr.

Fuel = [14.6 Btu/scf (Table II-1) X (8,000 scfm) X (\$2.00/million Btu*) X
(60 min/hr) X (8760 hr/yr) = \$122,780/yr.

Electricity =

$$\frac{(16 \text{ in. H}_2\text{O} \times (0.0001575 \frac{\text{hp}}{\text{in. H}_2\text{O}}) \times (8,000 \text{ scfm}) \times (1.45 \frac{\text{scf of air + off-gas}}{\text{scf of off-gas}})}{0.60 \text{ efficiency}} \times$$

$$\times (0.746 \text{ kWh}) \times (\$0.03/\text{kWh}) \times (8760 \text{ hr/yr}) = \$9,551/\text{yr}.$$

Labor = \$36,000/yr (from ref 1); the labor cost for a system without heat recovery is \$18,000/yr.

* See Table IV-2.

$$\text{Credit} = (8.56 \text{ Btu/scf}) \times (8,000 \text{ scfm}) \times \left(1.45 \frac{\text{scf of flue gas}}{\text{scf of waste gas}}\right) \times (60 \text{ min/hr}) \times \\ \times (\$2.00/\text{million Btu}) \times (8760 \text{ hr/yr}) = \$104,000/\text{yr}.$$

$$\text{Annual cost} = \text{fixed cost } (\$179,000/\text{yr}) + \text{operating cost } (\$168,000/\text{yr}) - \\ \text{credit } (\$104,000/\text{yr}) = \$243,000/\text{yr}.$$

$$\text{Net cost} = \frac{\text{annual cost } (\$243,000/\text{yr})}{\text{flow rate } (8,000 \text{ scfm})} = \$30.4/\text{scfm}.$$

ANNUAL COSTS OF CATALYTIC OXIDATION SYSTEMS

90% CONVERSION OF VOC
OFF-GAS HEAT CONTENT: 0.0 BTU/SCF

CASE	OFF-GAS FLOW (SCFH)	CAPITAL COST (\$1000)	FIXED COST	OPERATING COST OR CREDIT (\$1000)	RECOVERY CREDIT	NET ANNUALIZED COST OR CREDIT (\$1000)	NET COST OR SAVINGS (\$/SCFH)
NO HEAT RECOVERY	700.	184.	54.	36.	0.	90.	129.
	2000.	236.	71.	69.	0.	140.	70.
	5000.	314.	98.	145.	0.	243.	49.
	10000.	411.	132.	273.	0.	405.	41.
	20000.	574.	192.	528.	0.	720.	36.
	50000.	1022.	359.	1293.	0.	1652.	33.
RECUPERATIVE HEAT EXCHANGER 37% HEAT RECOVERY	100000.	1812.	649.	2568.	0.	3216.	32.
	700.	236.	69.	38.	0.	107.	154.
	2000.	311.	93.	59.	0.	151.	76.
	5000.	429.	131.	106.	0.	237.	47.
	10000.	582.	181.	185.	0.	366.	37.
	20000.	841.	268.	342.	0.	610.	31.
WASTE HEAT BOILER, 100 PSIG STEAM 50% HEAT RECOVERY	50000.	1547.	507.	815.	0.	1322.	26.
	100000.	2740.	910.	1602.	0.	2512.	25.
	700.	251.	74.	54.	9.	119.	170.
	2000.	322.	96.	88.	26.	138.	79.
	5000.	444.	135.	167.	65.	237.	47.
	10000.	612.	191.	298.	130.	338.	36.
50000.	20000.	909.	290.	559.	261.	588.	29.
	50000.	1716.	561.	1345.	652.	1253.	25.
	100000.	2950.	979.	2653.	1305.	2327.	23.

ANNUAL COSTS OF CATALYTIC OXIDATION SYSTEMS

 90% CONVERSION OF VOC
 OFF-GAS HEAT CONTENT: 2.0 BTU/SCF

CASE	OFF-GAS FLOW (SCFH)	CAPITAL COST (\$1000)	FIXED COST	OPERATING COST OR CREDIT (\$1000)	RECOVERY CREDIT	NET ANNUALIZED COST OR CREDIT (\$1000)	NET COST OR SAVINGS (\$/SCFH)
NO HEAT RECOVERY	700.	184.	54.	36.	0.	91.	129.
	2000.	237.	72.	70.	0.	141.	71.
	5000.	316.	98.	147.	0.	246.	49.
	10000.	415.	134.	276.	0.	410.	41.
	20000.	580.	195.	534.	0.	729.	36.
	50000.	1037.	365.	1308.	0.	1673.	33.
RECUPERATIVE HEAT EXCHANGER 37% HEAT RECOVERY	100000.	1845.	661.	2598.	0.	3259.	33.
	700.	236.	69.	38.	0.	108.	154.
	2000.	311.	93.	59.	0.	152.	76.
	5000.	429.	131.	107.	0.	237.	47.
	10000.	582.	181.	186.	0.	367.	37.
	20000.	841.	268.	345.	0.	613.	31.
WASTE HEAT BOILER, 100 PSIG STEAM 50% HEAT RECOVERY	50000.	1547.	507.	822.	0.	1329.	27.
	100000.	2740.	910.	1617.	0.	2527.	25.
	700.	252.	74.	55.	9.	119.	170.
	2000.	323.	97.	89.	27.	159.	79.
	5000.	446.	136.	168.	67.	238.	48.
	10000.	615.	192.	301.	133.	359.	36.
WASTE HEAT BOILER, 100 PSIG STEAM 50% HEAT RECOVERY	20000.	915.	292.	566.	266.	591.	30.
	50000.	1731.	566.	1360.	666.	1261.	25.
	100000.	2983.	991.	2685.	1332.	2344.	23.

ANNUAL COSTS OF CATALYTIC OXIDATION SYSTEMS

90% CONVERSION OF VOC
OFF-GAS HEAT CONTENT: 10.0 BTU/SCF

CASE	OFF-GAS FLOW (SCFM)	CAPITAL COST (\$1000)	FIXED COST	OPERATING COST OR CREDIT (\$1000)	RECOVERY CREDIT	NET ANNUALIZED COST OR CREDIT (\$1000)	NET COST OR SAVINGS (\$/SCFM)
NO HEAT RECOVERY	700.	184.	54.	29.	0.	83.	119.
	2000.	236.	71.	49.	0.	121.	60.
	5000.	314.	98.	97.	0.	194.	39.
	10000.	411.	132.	175.	0.	308.	31.
	20000.	574.	192.	333.	0.	525.	26.
	50000.	1022.	359.	805.	0.	1164.	23.
RECUPERATIVE HEAT EXCHANGER 36% HEAT RECOVERY	100000.	1812.	649.	1592.	0.	2241.	22.
	700.	235.	69.	32.	0.	101.	144.
	2000.	309.	92.	41.	0.	133.	66.
	5000.	426.	130.	61.	0.	191.	38.
	10000.	576.	179.	96.	0.	275.	28.
	20000.	830.	285.	165.	0.	429.	21.
WASTE HEAT BOILER, 100 PSIG STEAM 50% HEAT RECOVERY	50000.	1520.	500.	371.	0.	871.	17.
	100000.	2686.	894.	715.	0.	1609.	16.
	700.	251.	74.	47.	9.	112.	160.
	2000.	322.	96.	69.	26.	139.	69.
	5000.	444.	135.	118.	65.	188.	38.
	10000.	612.	191.	200.	130.	260.	26.
50000.	20000.	909.	290.	364.	261.	393.	20.
	50000.	1716.	561.	857.	652.	765.	15.
	100000.	2950.	979.	1678.	1305.	1351.	14.

ANNUAL COSTS OF CATALYTIC OXIDATION SYSTEMS

90% CONVERSION OF VOC
OFF-GAS HEAT CONTENT: 20.0 BTU/SCF

CASE	OFF-GAS FLOW (SCFM)	CAPITAL COST (\$1000)	OPERATING FIXED COST	OPERATING COST OR CREDIT (\$1000)	RECOVERY CREDIT	NET ANNUALIZED COST OR CREDIT (\$1000)	NET COST OR SAVINGS (\$/SCFM)
NO HEAT RECOVERY							
	700.	187.	55.	30.	0.	85.	122.
	2000.	241.	73.	52.	0.	125.	63.
	5000.	323.	101.	104.	0.	205.	41.
	10000.	426.	138.	189.	0.	327.	33.
	20000.	600.	202.	360.	0.	562.	28.
	50000.	1087.	383.	873.	0.	1257.	25.
	100000.	1958.	702.	1729.	0.	2430.	24.
RECUPERATIVE HEAT EXCHANGER							
45% HEAT RECOVERY							
	700.	244.	72.	28.	0.	99.	142.
	2000.	326.	97.	29.	0.	126.	63.
	5000.	459.	139.	32.	0.	171.	34.
	10000.	633.	196.	37.	0.	233.	23.
	20000.	933.	295.	47.	0.	342.	17.
	50000.	1763.	572.	77.	0.	649.	13.
	100000.	3182.	1041.	127.	0.	1168.	12.
WASTE HEAT BOILER, 100 PSIG STEAM							
62% HEAT RECOVERY							
	700.	258.	76.	48.	16.	108.	155.
	2000.	338.	101.	72.	46.	126.	63.
	5000.	478.	146.	125.	116.	156.	31.
	10000.	676.	210.	214.	231.	194.	19.
	20000.	1027.	326.	393.	462.	257.	13.
	50000.	1962.	637.	928.	1155.	410.	8.
	100000.	3295.	1090.	1820.	2310.	599.	6.

ANNUAL COSTS OF CATALYTIC OXIDATION SYSTEMS

90% CONVERSION OF VOC

OFF-GAS HEAT CONTENT: 10.0 BTU/SCF (IN AIR)

CASE	OFF-GAS FLOW (SCFM)	CAPITAL COST (\$1000)	FIXED COST	OPERATING COST OR CREDIT (\$1000) OPERATING COST	RECOVERY CREDIT	NET ANNUALIZED COST OR CREDIT (\$1000)	NET COST OR SAVINGS (\$/SCFM)
NO HEAT RECOVERY	700.	172.	51.	26.	0.	76.	109.
	2000.	215.	64.	40.	0.	104.	52.
	5000.	279.	86.	72.	0.	158.	32.
	10000.	357.	113.	127.	0.	240.	24.
	20000.	482.	158.	236.	0.	394.	20.
RECUPERATIVE HEAT EXCHANGER 53% HEAT RECOVERY	50000.	802.	276.	563.	0.	839.	17.
	100000.	1330.	472.	1108.	0.	1579.	16.
	700.	233.	68.	28.	0.	96.	137.
	2000.	308.	91.	28.	0.	120.	60.
	5000.	428.	129.	31.	0.	160.	32.
WASTE HEAT BOILER, 100 PSIG STEAM 56% HEAT RECOVERY	10000.	585.	179.	34.	0.	213.	21.
	20000.	852.	265.	41.	0.	307.	15.
	50000.	1373.	500.	63.	0.	563.	11.
	100000.	2766.	888.	99.	0.	987.	10.
	700.	238.	70.	44.	6.	106.	151.
	2000.	299.	88.	59.	23.	124.	62.
	5000.	401.	121.	93.	58.	156.	31.
	10000.	542.	166.	150.	117.	199.	20.
	20000.	789.	247.	263.	233.	277.	14.
	50000.	1438.	461.	604.	583.	481.	10.
100000.	2390.	779.	1172.	1167.	1167.	784.	8.

ANNUAL COSTS OF CATALYTIC OXIDATION SYSTEMS

99% CONVERSION OF VNC
OFF-GAS HEAT CONTENT: 0.0 BTU/SCF

CASE	OFF-GAS FLOW (SCFM)	CAPITAL COST (\$1000)	OPERATING FIXED COST	OPERATING COST OR CREDIT (\$1000) RECOVERY CREDIT	NET ANNUALIZED COST OR CREDIT (\$1000)	NET COST OR SAVINGS (\$/SCFM)
NO HEAT RECOVERY	700.	185.	55.	36.	91.	130.
	2000.	240.	74.	69.	143.	72.
	5000.	324.	104.	146.	250.	50.
	10000.	431.	145.	274.	119.	42.
	20000.	613.	217.	531.	748.	37.
RECUPERATIVE HEAT EXCHANGER 37% HEAT RECOVERY	700.	1117.	418.	1300.	1719.	34.
	20000.	1996.	763.	2583.	3346.	33.
	50000.					
	100000.					
WASTE HEAT BOILER: 100 PSIG STEAM 50% HEAT RECOVERY	700.	238.	70.	38.	108.	155.
	2000.	315.	95.	39.	154.	77.
	5000.	439.	137.	106.	243.	49.
	10000.	601.	193.	186.	379.	38.
	20000.	878.	291.	345.	636.	32.
WASTE HEAT BOILER: 100 PSIG STEAM 50% HEAT RECOVERY	700.	1635.	562.	822.	1384.	28.
	2000.	2912.	1017.	1616.	2633.	26.
	5000.					
	10000.					
WASTE HEAT BOILER: 100 PSIG STEAM 50% HEAT RECOVERY	700.	253.	75.	54.	120.	172.
	2000.	326.	99.	89.	161.	81.
	5000.	454.	142.	168.	244.	49.
	10000.	632.	203.	299.	372.	37.
	20000.	948.	314.	562.	615.	31.
WASTE HEAT BOILER: 100 PSIG STEAM 50% HEAT RECOVERY	50000.	1811.	619.	1352.	1319.	26.
	100000.	3134.	1093.	2668.	2457.	25.

ANNUAL COSTS OF CATALYTIC OXIDATION SYSTEMS

99% CONVERSION OF VOC
OFF-GAS HEAT CONTENT: 2.0 BTU/SCF

CASE	OFF-GAS FLOW (SCFM)	CAPITAL COST (\$1000)	FIXED COST	OPERATING COST OR CREDIT (\$1000)	RECOVERY CREDIT	NET ANNUALIZED COST OR CREDIT (\$1000)	NET COST OR SAVINGS (\$/SCFM)
NO HEAT RECOVERY	700.	186.	55.	36.	0.	92.	131.
	2000.	241.	74.	70.	0.	144.	72.
	5000.	327.	105.	148.	0.	253.	51.
	10000.	435.	146.	278.	0.	424.	42.
	20000.	620.	219.	537.	0.	756.	38.
RECUPERATIVE HEAT EXCHANGER 37% HEAT RECOVERY	50000.	1133.	425.	1316.	0.	1741.	35.
	100000.	2033.	778.	2613.	0.	3391.	34.
	700.	238.	70.	38.	0.	109.	155.
	2000.	315.	95.	59.	0.	154.	77.
	5000.	439.	137.	107.	0.	244.	49.
WASTE HEAT BOILER, 100 PSIG STEAM 50% HEAT RECOVERY	10000.	601.	193.	187.	0.	380.	38.
	20000.	878.	291.	348.	0.	639.	32.
	50000.	1635.	562.	829.	0.	1391.	28.
	100000.	2912.	1017.	1631.	0.	2648.	26.
	700.	253.	75.	55.	9.	120.	172.
50% HEAT RECOVERY	2000.	328.	99.	89.	27.	162.	81.
	5000.	456.	143.	169.	67.	245.	49.
	10000.	635.	205.	302.	133.	374.	37.
	20000.	955.	317.	569.	266.	619.	31.
	50000.	1828.	626.	1348.	666.	1328.	27.
	100000.	3171.	1108.	2700.	1332.	2476.	25.

ANNUAL COSTS OF CATALYTIC OXIDATION SYSTEMS

99% CONVERSION OF VOC
OFF-GAS HEAT CONTENT: 10.0 BTU/SCF

CASE	OFF-GAS FLOW (SCFH)	CAPITAL COST (\$1000)	OPERATING COST OR CREDIT (\$1000) FIXED COST OPERATING COST	RECOVERY CREDIT	NET ANNUALIZED COST OR CREDIT (\$1000)	NET COST OR SAVINGS (\$/SCFH)
NO HEAT RECOVERY						
	700.	185.	55.	0.	84.	120.
	2000.	240.	74.	0.	124.	62.
	5000.	324.	104.	0.	202.	40.
	10000.	431.	145.	0.	322.	32.
	20000.	613.	217.	0.	553.	28.
	50000.	1117.	418.	0.	1231.	25.
	100000.	1996.	763.	0.	2371.	24.
RECUPERATIVE HEAT EXCHANGER						
36% HEAT RECOVERY	700.	237.	70.	0.	102.	146.
	2000.	313.	95.	0.	136.	68.
	5000.	435.	136.	0.	198.	40.
	10000.	594.	191.	0.	288.	29.
	20000.	866.	287.	0.	455.	23.
	50000.	1608.	554.	0.	933.	19.
	100000.	2858.	1001.	0.	1730.	17.
WASTE HEAT BOILER, 100 PSIG STEAM						
50% HEAT RECOVERY	700.	253.	75.	9.	113.	162.
	2000.	326.	99.	26.	142.	71.
	5000.	454.	142.	65.	195.	39.
	10000.	632.	203.	130.	274.	27.
	20000.	948.	314.	261.	420.	21.
	50000.	1811.	619.	652.	831.	17.
	100000.	3134.	1093.	1305.	1481.	15.

ANNUAL COSTS OF CATALYTIC OXIDATION SYSTEMS

99% CONVERSION OF VHC
OFF-GAS HEAT CONTENT: 20.0 BTU/SCF

CASE	OFF-GAS FLOW (SCFM)	CAPITAL COST (\$1000)	FIXED COST	OPERATING COST OR CREDIT (\$1000) RECOVERY CREDIT	ANNUALIZED COST OR CREDIT (\$1000)	NET COST OR SAVINGS (\$/SCFM)
NO HEAT RECOVERY	700.	188.	56.	30.	86.	123.
	2000.	246.	76.	53.	128.	64.
	5000.	334.	108.	104.	212.	42.
	10000.	448.	151.	191.	342.	34.
	20000.	643.	229.	363.	592.	30.
RECUPERATIVE HEAT EXCHANGER 45% HEAT RECOVERY	50000.	1189.	447.	882.	1329.	27.
	100000.	2158.	826.	1745.	2571.	26.
	700.	245.	73.	28.	100.	143.
	2000.	330.	100.	29.	129.	65.
	5000.	468.	146.	33.	178.	36.
WASTE HEAT BOILER, 100 PSIG STEAM 62% HEAT RECOVERY	10000.	652.	208.	38.	247.	25.
	20000.	870.	319.	50.	369.	18.
	50000.	1854.	628.	84.	712.	14.
	100000.	3358.	1151.	141.	1293.	13.
	700.	260.	77.	49.	109.	156.
	2000.	343.	104.	72.	130.	65.
	5000.	489.	153.	126.	163.	33.
	10000.	697.	224.	231.	209.	21.
	20000.	1070.	352.	394.	286.	14.
	50000.	2064.	701.	715.	482.	10.
	100000.	3495.	1214.	1836.	740.	7.

ANNUAL COSTS OF CATALYTIC OXIDATION SYSTEMS

99% CONVERSION OF VOC
OFF-GAS HEAT CONTENT: 10.0 BTU/SCF (IN AIR)

CASE	OFF-GAS FLOW (SCFH)	CAPITAL COST (\$1000)	FIXED COST	OPERATING COST OF CREDIT (\$1000) OPERATING COST	RECOVERY CREDIT	NET ANNUALIZED COST OR CREDIT (\$1000)	NET COST OR SAVINGS (\$/SCFH)
NO HEAT RECOVERY	700.	173.	51.	26.	0.	77.	110.
	2000.	218.	66.	40.	0.	106.	53.
	5000.	287.	90.	73.	0.	163.	33.
	10000.	371.	121.	128.	0.	249.	25.
	20000.	509.	175.	238.	0.	413.	21.
RECUPERATIVE HEAT EXCHANGER 53% HEAT RECOVERY	50000.	868.	318.	568.	0.	886.	19.
	100000.	1458.	552.	1118.	0.	1670.	17.
	700.	234.	69.	28.	0.	97.	138.
	2000.	311.	93.	29.	0.	122.	61.
	5000.	436.	134.	31.	0.	163.	33.
WASTE HEAT BOILER, 100 PSIG STEAM 56% HEAT RECOVERY	10000.	599.	188.	35.	0.	223.	22.
	20000.	880.	282.	43.	0.	326.	16.
	50000.	1639.	541.	68.	0.	609.	12.
	100000.	2895.	968.	109.	0.	1078.	11.
	700.	240.	71.	44.	8.	106.	152.
	2000.	302.	90.	59.	23.	126.	63.
	5000.	409.	126.	93.	58.	161.	32.
	10000.	556.	175.	151.	117.	209.	21.
	20000.	816.	264.	265.	233.	286.	15.
	50000.	1504.	502.	583.	583.	609.	11.
	100000.	2519.	859.	1182.	1167.	873.	9.

REPORT 4

CONTROL DEVICE EVALUATION
FLARES AND THE USE OF EMISSIONS AS FUELS

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Prepared for
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Research Triangle Park, North Carolina

February 1981

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I. INTRODUCTION

This control device evaluation deals with volatile organic compound emissions (VOC) being destroyed by a high-temperature oxidation flame to the normal combustion products carbon dioxide and water. Three types of systems are evaluated: an elevated flare, an enclosed ground flare, and using the emissions as a fuel. To be controlled by these systems the emission must have sufficient fuel value, or be enriched with auxiliary fuel, for a stable flame to be maintained. Normal safety precautions for handling combustible gases must be exercised. If the emission contains constituents such as sulfur and halogen compounds, they will be oxidized to noxious gases that will be discharged to the atmosphere.

The elevated flare, which is a single burner tip elevated above ground level for safety reasons, burns the vented gases in essentially a diffusion flame. This type of system, especially in the large sizes, can create problems with luminosity, combustion noise, and heat radiation.

The ground flare is composed of multiple gas burner heads that are grouped in an enclosure and are staged to operate based on the flow of the vented gas. The enclosures reduce the luminosity, noise, and radiation problems and allows the flare to be located at ground elevation.

For the emissions to be used as a fuel the need for it must be established and transportation from the point of generation to the point of use must be possible. These restrictions usually limit this control method to emissions that have a fairly steady generation characteristic.

The three control systems can be used in combinations for improved handling of emissions, for instance, for a facility that has some steady emissions, has potential for some fairly frequent intermediate-size emissions, and requires infrequent emergency venting. The fuel-gas system could be designed to handle the steady emissions; emissions in excess of the fuel needed would go to an enclosed ground flare sized to handle the intermediate-size emissions; then further excess emissions would go to an elevated flare sized to handle the emergency

venting. This combined system would maximize energy conservation, burn most of the emissions in a potentially more efficient and acceptable manner, and yet be capable of safely handling emergency conditions. Because a combined system is very site-specific this report treats each control method independently.

II. SYSTEM DESCRIPTIONS

A. ELEVATED FLARES

There are three general types of elevated flares: those that are nonsmokeless, those that are smokeless, and those that are fired or endothermic. The first type, which is the most simple one, is just a flare tip without any special provision for enhancing the mixing of air into its flame. Its use is limited essentially to gas streams that burn readily without producing smoke, for example, streams that contain predominately methane, hydrogen, or carbon monoxide. In the second type, which is the one most widely used and is the most versatile, an exterior momentum source is used to improve the flame-air mixing and turbulence. The momentum sources that have been used are steam, water sprays, high-pressure gas, air blowers, and compressed air. The third type uses a high-energy-content fuel gas to provide heating value to a lean flare gas that will not support a stable flame by itself, usually a gas with less than 115 Btu/ft³ heating value.^{1*} This type of flare has very limited use. It has been used for disposal of such streams as sulfur tail gases and ammonia waste gases. Although each flare type may have potential application to a specific situation in the synthetic organic chemicals manufacturing industry, this study deals only with the steam-assisted smokeless flare, which has been reported to account for greater than 95% of the flares installed.²

Figure II-1 is a diagram of a steam-assisted elevated smokeless flare system showing the usual components that are used. The emission source gas is conveyed by a transfer line from the facility release point to the flare location. The line is equipped for purging so that explosive mixtures do not occur in the flare system either on startup or during operation. The usual purge gas is natural gas, although other fuel gases and inert gases such as nitrogen can be used.

Liquids that may be in the emission source gas or that may condense out in the collection headers and transfer line are removed by a disentrainment drum located close to the flare. Liquids in a flare gas can cause smoke to form because of incomplete burning and if the size of the droplets³ is greater than

*See Sect. VII for the references cited in this report.

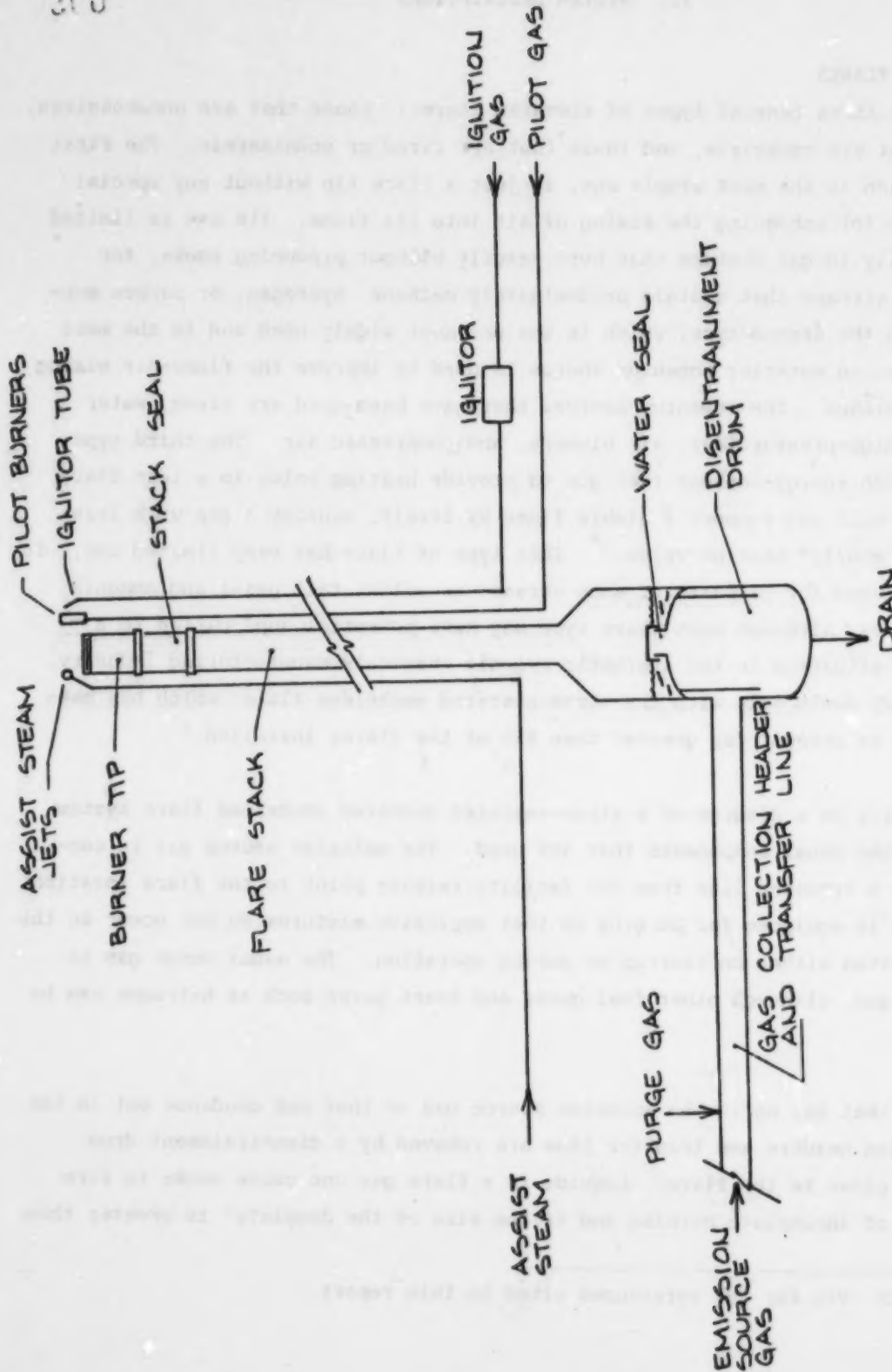


Fig. II-1. Diagram of Steam-Assisted Smokeless Elevated-Flare System

150 μm may generate a spray of burning chemicals that could reach ground level and present a safety hazard. A water seal is usually located between the dis-entrainment drum and the flare stack to prevent flashbacks into the system. Other devices, such as flame arresters and actuated check valves, may sometimes replace a water seal or be used in conjunction with it.

For safety reasons a stack is used to elevate the flare. The flare must be located so that it does not present a hazard to surrounding personnel and facilities; therefore the usual choice is to elevate it. The height is determined by designing for a maximum ground-level heat radiation and possibly also for plume dispersion in case of emission ignition failure. A stack seal is normally used just below the flare tip to impede the incursion of air back into the flare system, which could create an explosion potential. The use of a seal reduces the operating purge-gas requirements.

The burner tip is designed to give an environmentally acceptable combustion of the flare gas over the flare system's capacity range. Consideration is given to flame stability, ignition reliability, effective assist-steam injection, and, possibly, noise suppression. The burner tips are normally proprietary in design. Flame stability can be enhanced by flame holder retention devices incorporated in the flare tip inner circumference. With burner tips with modern flame holder designs the flame can be stable over a flare-gas exit discharge velocity range of 1 to 600 fps.¹ Reliable ignition is obtained by continuous pilot burners designed for stability and positioned around the outer perimeter of the flare tip. The number of pilot burners required depends on flare size and, possibly, on flare gas composition and wind conditions. The pilot burners are ignited by an ignition source system, which can be designed for either manual or automatic actuation. Effective assist-steam injection to promote turbulence and mixing of air into the flare gas flame for more efficient combustion and smoke suppressions is generally accomplished by using high-velocity steam jets positioned around the outer perimeter of the flare tip. For the larger flares steam can also be injected concentrically into the flare tip. For some proprietary designs steam injection includes the aspiration of air along with the steam.

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Steam flow can be controlled manually but automatic control, based on flare gas flow and flame radiation, gives a faster response to the need for steam and a better adjustment of the quantity required. The physical limitation on the quantity of steam that can be delivered and injected into the flare flame determines the smokeless capacity of the flare, which is usually less than its stable flame capacity. The use of steam injection into a flare flame can produce other results in addition to air entrainment and turbulence. For example, the water in the steam can enter into a water/gas reaction with carbon or into a steam—re-forming reaction with hydrocarbons; either reaction can reduce smoke formation. Steam can moderate the flame temperature, which could inhibit flare-gas constituents from participating in cracking reactions that form carbon and smoke. A detrimental effect of steam usage is that it can aggravate the flare noise problem by producing high-frequency jet noise and by increasing the combustion rate. The jet noise can be reduced by the use of small multiple steam jets and, if necessary, by accoustical shrouding.

B. ENCLOSED GROUND FLARES

An enclosed ground flare has multiple burner heads that are staged to operate based on the quantity of gas being released to the flare. The size, design, number, and arrangement of the burner heads depend on the flare-gas characteristics. An exterior motivation source, such as steam or air, to enhance combustion and prevent smoke formation is not required except in rare applications. Stable combustion can be obtained with some gases that have heat contents as low as 50 to 60 Btu/ft³ (ref 4). Reliable and efficient operation can be attained from 0 to 100% of capacity.⁴

Figure II-2 is a diagram of an enclosed ground flare system with the components that are normally used. The emission source gas is conveyed from its facility release point to the flare location by a transfer line. Purge gas probably is needed only for initial purging of the system on startup. Liquids that may be in the emission source gas or that may condense out in the collection headers and transfer line are removed by a disentrainment drum located close to the flare. If a potential for overloading the ground-flare system exists, the excess gas may need to be diverted to an elevated flare or other safety precautions taken. A water seal can be located between the disentrainment drum and the ground flare if the potential exists for flashbacks into the system.

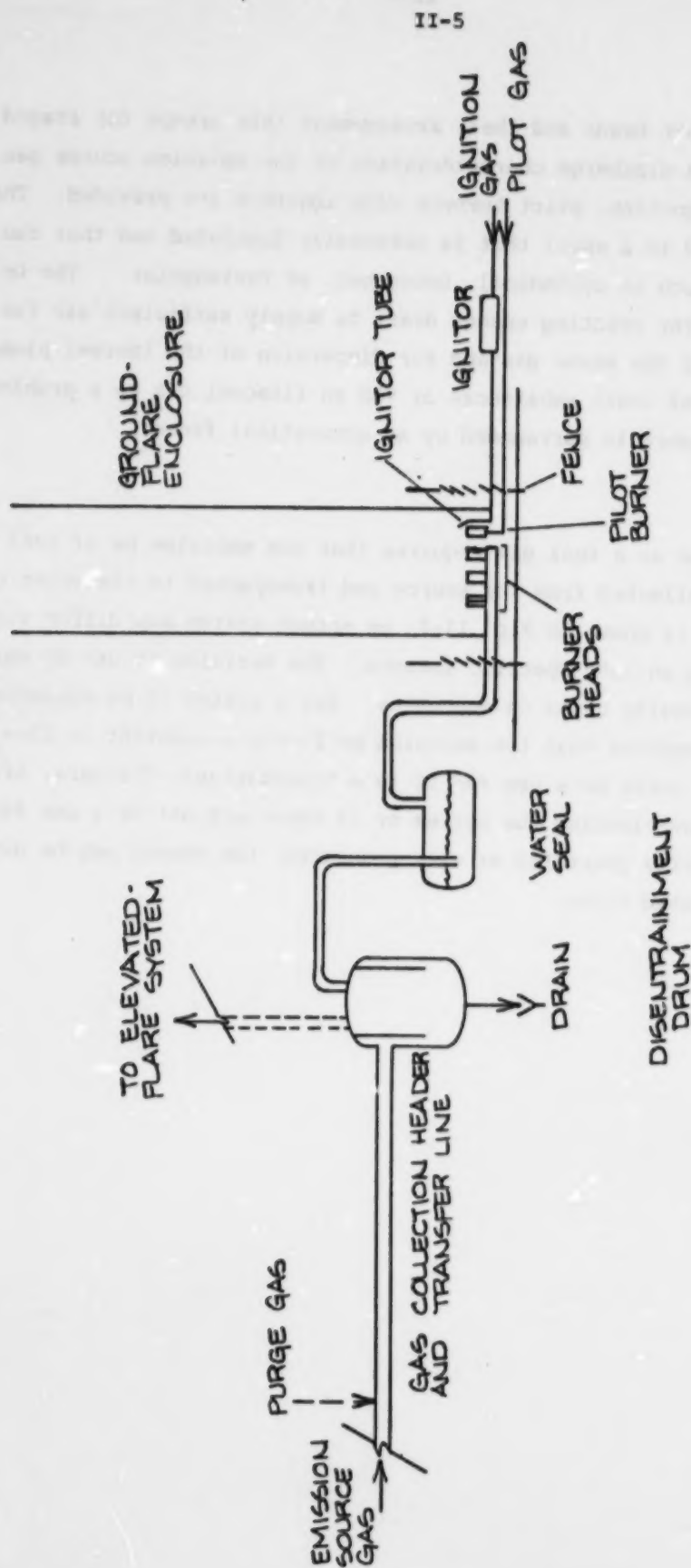


Fig. II-2. Diagram of Enclosed Ground-Flare System

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The number of burner heads and their arrangement into groups for staged operation depend on the discharge characteristics of the emission source gas. To ensure reliable ignition, pilot burners with ignitors are provided. The burner heads are enclosed in a shell that is internally insulated and that can be of several shapes, such as cylindrical, hexagonal, or rectangular. The height must be adequate for creating enough draft to supply sufficient air for smokeless combustion of the waste gas and for dispersion of the thermal plume. Also, dispersion of toxic substances or VOC on flameout can be a problem. The base of the enclosure is surrounded by an accoustical fence.

C. FUEL-GAS SOURCE

To use an emission as a fuel gas requires that the emission be of fuel quality and that it be collected from its source and transported to the point of use. A typical system is shown in Fig. II-3; an actual system may differ significantly, depending on site-specific factors. The decision to use an emission as a fuel gas is normally based on economics. For a system to be economically viable usually requires that the emission be fairly consistent in flow and quality and that there be a use for it in a transportable distance. If there is a potential for overloading the system or if there may not be a use for the fuel in the quantity generated or when generated, the excess may be diverted to a ground or elevated flare.

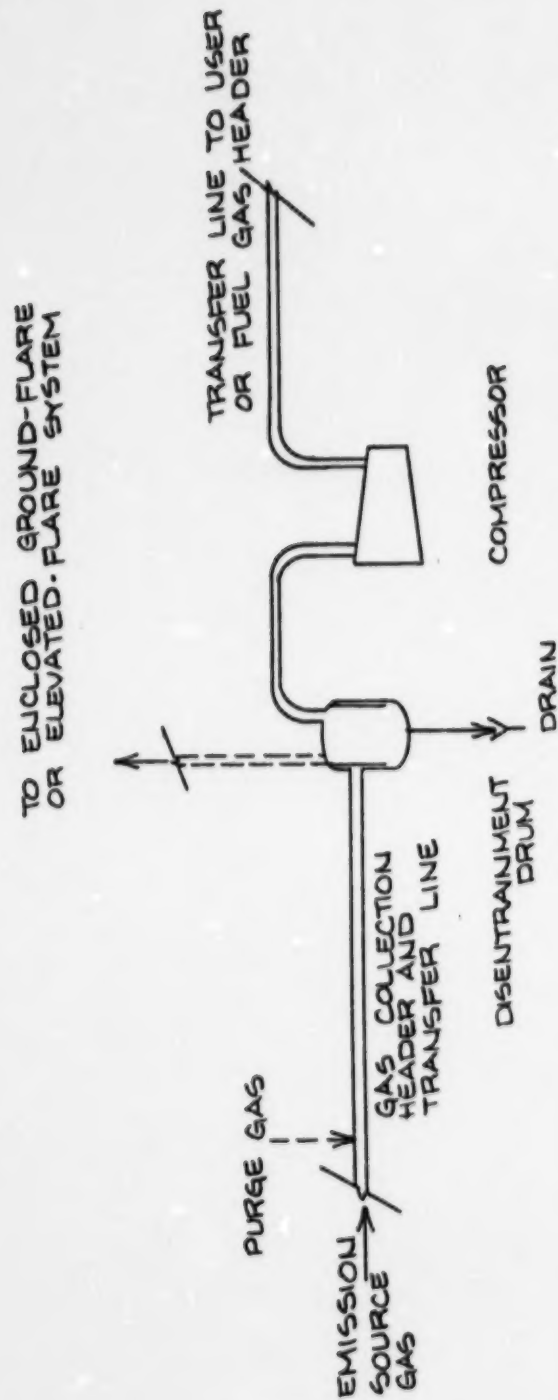


Fig. II-3. Diagram of Emission-Fuel-Gas System

III. SYSTEM EFFICIENCIES

Little information is presently available on the combustion efficiency of flares. Although some test data are available (and discussed below), the results are not applicable to the combustion of typical SOCHI off-gas streams in typical SOCHI flares.

Flare efficiencies of 98% and 99% are used for the purpose of cost-effectiveness calculations in this report. It should not be assumed that typical SOCHI flares will obtain efficiencies in this range. Well-designed and well-operated small flares burning easily combusted gases do obtain high efficiencies but further work is necessary to determine efficiencies of typical flares in this industry. The EPA is presently (1981) conducting studies on the efficiencies obtained by a variety of flares at typical operating conditions.⁵

A. STEAM-ASSISTED ELEVATED FLARES

An elevated flare without steam assistance would burn as one large diffusion flame. A diffusion flame is one in which air diffuses across the boundary of the fuel/combustion product stream toward the center of the fuel flow, forming an envelope of a combustible gas mixture around a core of fuel gas. This stream, on ignition, establishes a stable flame zone around a gas core above the burner tip. This inner gas core is heated by diffusion of hot combustion products from the flame zone. Cracking can occur with the formation of small hot particles of carbon that give the flame its characteristic luminosity. If there is an oxygen deficiency and if the carbon particles are cooled to below their ignition temperature, they can escape the flame zone as smoke. Also, in large diffusion flames, combustion product vortices can form around burning portions of the gas and shut off the supply of oxygen. This localized instability causes elongations and contractions (flame flickering), which can be accompanied by soot formation. Steam jets directed into the flame or used to aspirate air into the flowing gas improve the mixing of air with the fuel and give the flame some premixed flame character, thereby reducing or eliminating soot formation.

Sampling a commercial-size flare is very difficult with the methods now available and results obtained from testing small flares cannot at the present time

be scaled up for commercial-size flares. Three programs are reported for testing smaller sized commercial flares.^{1,6,7} In one program flares with 2-in., 3-in., and 6-in. tip sizes were tested;¹ results are given for selected tests with natural gas as the fuel and without the use of assist steam. From those results the methane hydrocarbon feed destruction efficiency was calculated to be >90% in only two tests and <70% in four tests. In another program⁶ the flare tip size was 16 in.; the flare gas was a mercaptan in nitrogen and assist steam was not used. The efficiencies reported ranged from 92 to >99%.

An extensive flare test program has been conducted in Germany.⁷ The flare tip was a truncated cone 20 cm (7.9 in.) at the flare-gas entrance, 50 cm (19.7 in.) long, and 70 cm (27.6 in.) at the exit. The cone was fitted with six steam nozzles that aspirated air into the flaming gas before it exited the cone. The operating parameters used were the following: flare gas flow, 0.13 to 2.9 mt/hr; gas composition density, 0.54 to 1.86 kg/m³; steam-to-gas weight ratio, 0 to 1.73; and cross wind, 0 to 6 m/s. The gases combusted had a high proportion of hydrogen and a low proportion of unsaturated hydrocarbons. Typically the hydrogen concentration was greater than 50%. Of 1298 measurements at the flame end and downstream from the flame end for the complete range of test conditions, in only 4 measurements were the local burnouts to carbon dioxide less than 99%.

The efficiencies used for cost-effectiveness calculations in this report are 99% for flares to 12 in. in diameter and 98% for flares over 12 in. in diameter.

B. ENCLOSED GROUND FLARES

An enclosed ground flare burns with multiple small diffusion flames from burner heads that can be stage-operated based on flare gas flow and that are enclosed on the sides. An enclosed ground flare has the control capabilities to maximize combustion efficiency for most operating conditions. The burner heads can be sized and designed for the materials in the flare gas; they can be stage-operated to give a turndown from 100 to 0%⁴ of design; and the enclosure design allows for a degree of combustion air and temperature control.

Although the design of an enclosed ground flare would allow the combustion gases to be monitored to determine combustion efficiency, there are no reported test programs on commercial-sized units. One reference¹ cites that a performance estimate was made for an open-flare flame.

The efficiency used in this report for cost-effectiveness calculations is 99%.

C. FUEL-GAS SOURCE

When an emission is used as a fuel gas, the VOC destruction efficiency depends on the waste gas composition and on the combustion conditions. The efficiency used in this report for cost-effectiveness calculations is >99.9%. According to the emission factors given in AP-42⁸ for natural gas and LPG when burned under proper operating conditions, this efficiency is attained. This combustion efficiency does not apply to dilute VOC streams introduced into a combustion device in any way other than as a fuel gas.

IV. DESIGN BASIS

The capital and operating costs developed for this study are based on model designs discussed in this section. The three control systems, steam-assisted elevated flares, enclosed ground flares, and emissions used as a fuel source, are evaluated independently. For all systems a model-waste-gas emission was used that had properties equivalent to those of propylene. The standard construction material is carbon steel except where it is standard practice to use other materials, such as burner tips.

A. STEAM-ASSISTED ELEVATED FLARES

1. Flare-Tip Capacity

The maximum and minimum capacity of a flare tip to burn a flared gas with a stable flame (not necessarily smokeless) is a function of tip design. At too high an exit velocity the flame can lift off the tip and flame out, while at too low a velocity it can burn back into the tip or lick down the sides of the stack. Modern commercial flares with flame retention rings are reported to have stable flame capabilities over flare-gas discharge velocities of 1 to 600 fps.¹ The actual maximum capacity of a flare tip is usually limited by the flare-gas pressure available to overcome the system pressure drop. For the purpose of determining flare heights the practical capacity in this study was assumed to be the model-waste-gas flow at ambient conditions, which results in a flare-tip pressure drop of 18 in. H₂O. For the larger flare-tip sizes these conditions would result in a discharge velocity of about 200 fps. Flare capacity based on the correlations in ref 9 (see Appendix A) is plotted in Fig. IV-1.

The capacity of a steam-assisted flare to burn smokelessly is limited by the quantity of steam that is available. There is a physical limit to the quantity of steam that can be effectively delivered to a flare tip. Figure IV-2 is a smoothed curve of the quantity that can be delivered based on a practical design of steam piping on the flare and on 150-lb steam pressure being available to the flare site.¹⁰ Another steam limit could be the availability of steam-generation capacity. In this study it was assumed that steam would be available to the flare site, on demand when needed.

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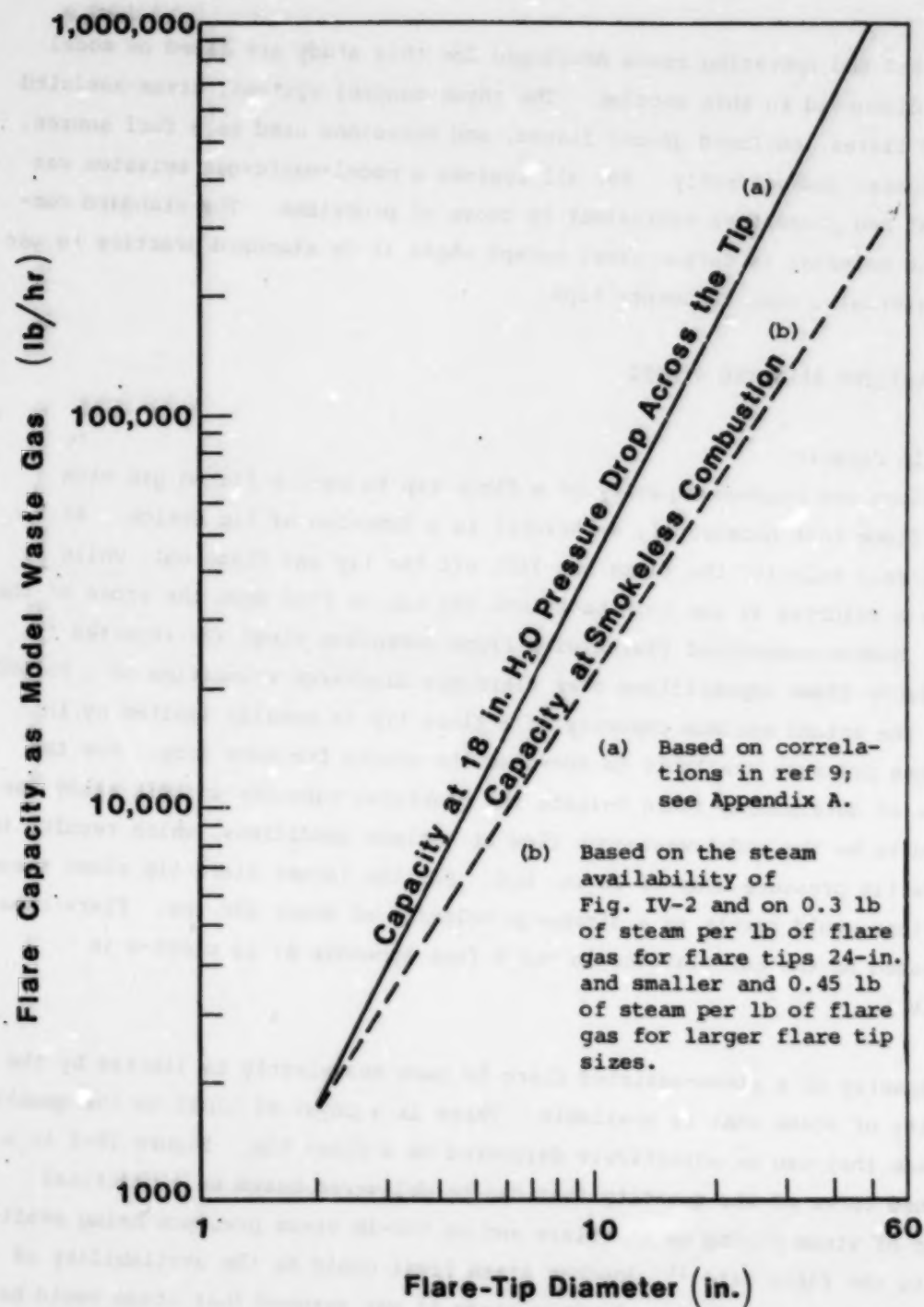


Fig. IV-1. Elevated-Flare Capacity

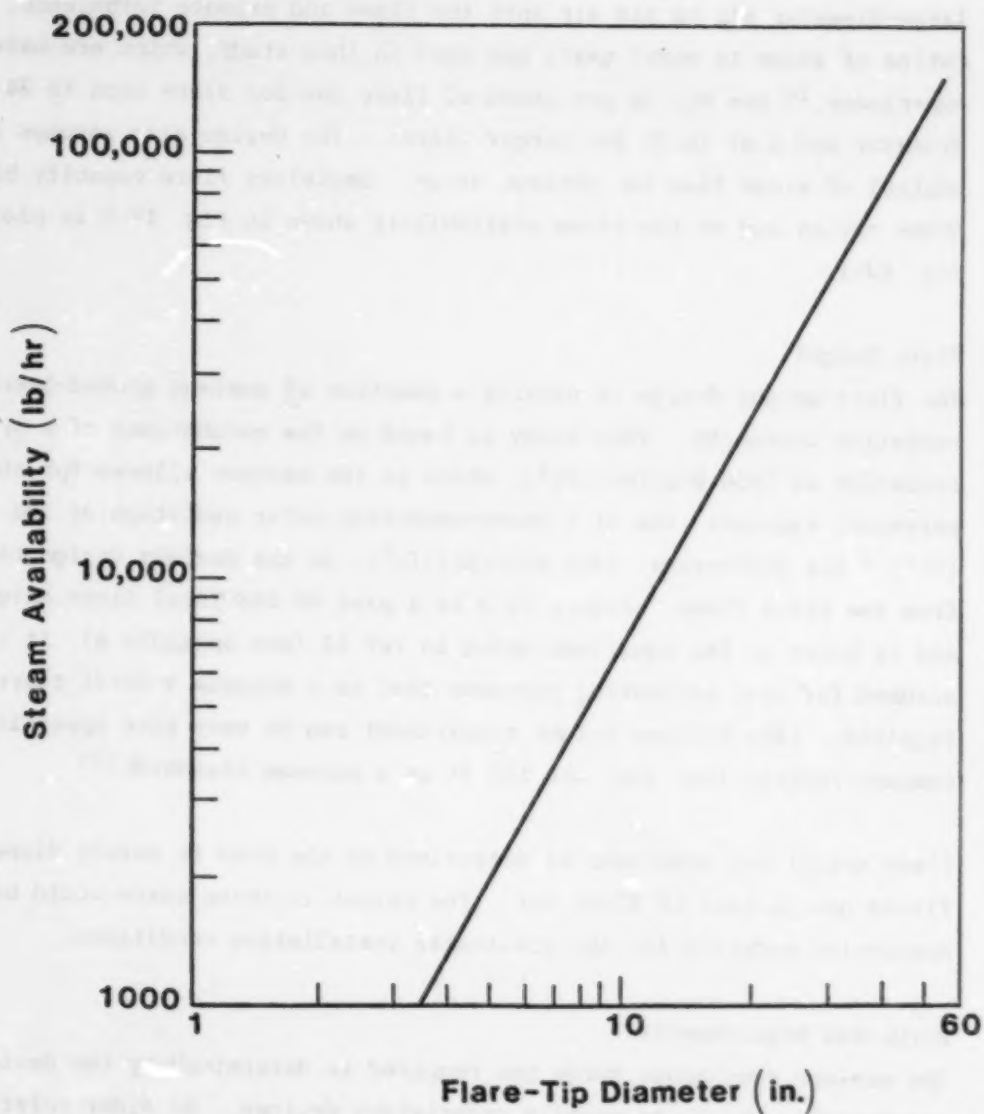


Fig. IV-2. Maximum Steam Availability at Elevated-Flare Tip

The steam requirement per quantity of flare gas depends on the composition of the gas flared and on the flare-tip design. Typical values range from 0.15 to 0.50 lb of steam per pound of flare gas.¹¹ Olefins, such as propylene, require higher steam ratios than would a paraffin hydrocarbon to burn smokelessly. A flare with a small-diameter tip can use steam more efficiently than a large-diameter tip to mix air into the flame and promote turbulence. The ratios of steam to model waste gas used in this study, which are based on experience,¹⁰ are 0.3 lb per pound of flare gas for flare tips to 24 in. in diameter and 0.45 lb/lb for larger flares. The design also assumes instrument control of steam flow for optimum usage. Smokeless flare capacity based on those ratios and on the steam availability shown in Fig. IV-2 is plotted in Fig. IV-1.

2. Flare Height

The flare-height design is usually a function of maximum ground-level heat-radiation intensity. This study is based on the assumptions of a ground-level radiation of 1500 Btu/(hr)(ft²), which is the maximum allowed for short-term personnel exposure, and of a sunny-condition solar radiation of 300 Btu/(hr)(ft²).³ The difference, 1200 Btu/(hr)(ft²), is the maximum design radiation from the flare flame. Figure IV-3 is a plot of the total flare heights used and is based on the equations given in ref 12 (see Appendix A). It is further assumed for cost estimating purposes that as a minimum a 40-ft flare height is required. (The minimum height requirement can be very site specific; one company reports that they use 150 ft as a minimum standard.)¹³

Flare height may sometimes be determined by the need to safely disperse the flared gas in case of flame out. The height in these cases would be based on dispersion modeling for the particular installation conditions.

3. Purge-Gas Requirements

The minimum continuous purge gas required is determined by the design of the stack seals, which are usually proprietary devices. An older reference¹⁴ reports that the gas velocity required is 0.10 to 0.15 fps; modern seals are stated to require less. This study is based on a flow of 0.03 fps,¹⁰ which is stated to be adequate for normal applications without any type of conservation. Use of various purge-gas conservation systems can reduce this consumption.¹⁵

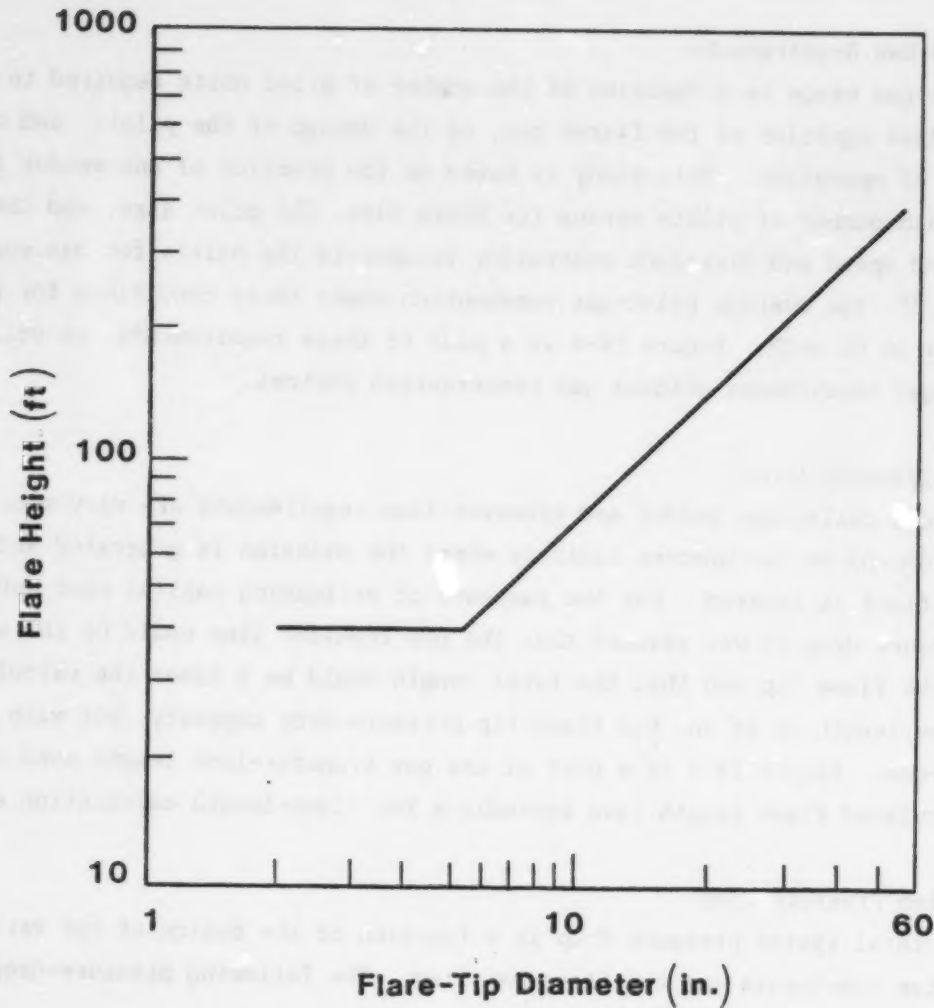


Fig. IV-3. Elevated-Flare Height

The purge gas is assumed to be a constant flow of natural gas at 60°F and 1 atm. Figure IV-4 shows a plot of these requirements. Purge gas also may be needed to purge the system before startup and to prevent a vacuum from sucking air back into the system after a hot gas discharge is flared. These uses are assumed to be minor.

4. Pilot-Gas Requirements

Pilot-gas usage is a function of the number of pilot units required to ensure positive ignition of the flared gas, of the design of the pilots, and of the mode of operation. This study is based on the practice of one vendor for a certain number of pilots versus the flare size, the pilot size, and the use of a wind speed and direction controller to operate the pilots for gas conservation.¹⁶ The average pilot-gas consumption under these conditions for all flare sizes is 60 scfh. Figure IV-4 is a plot of these requirements, as well as of the gas requirement without gas conservation control.

5. Gas Transfer Lines

The gas collection header and transfer-line requirements are very site specific and depend on the process facility where the emission is generated and on where the flare is located. For the purposes of estimating capital cost and system pressure drop it was assumed that the gas transfer line would be the same size as the flare tip and that the total length would be 3 times the calculated flame length at 18 in. H₂O flare-tip pressure-drop capacity, but with a 100-ft minimum. Figure IV-5 is a plot of the gas transfer-line length used and the calculated flame length (see Appendix A for flame-length calculation equation).

6. System Pressure Drop

The total system pressure drop is a function of the design of the various system components and the flare-gas flow. The following pressure-drop relationships are assumed:^{10,16}

Flare tip, calculated (see Appendix A)

Stack seal, 1.5 times flare-tip ΔP

Stack

Water seal

Disentrainment drum

} 0.5 times flare-tip ΔP

Transfer line, calculated based on diameter, length, and gas flow

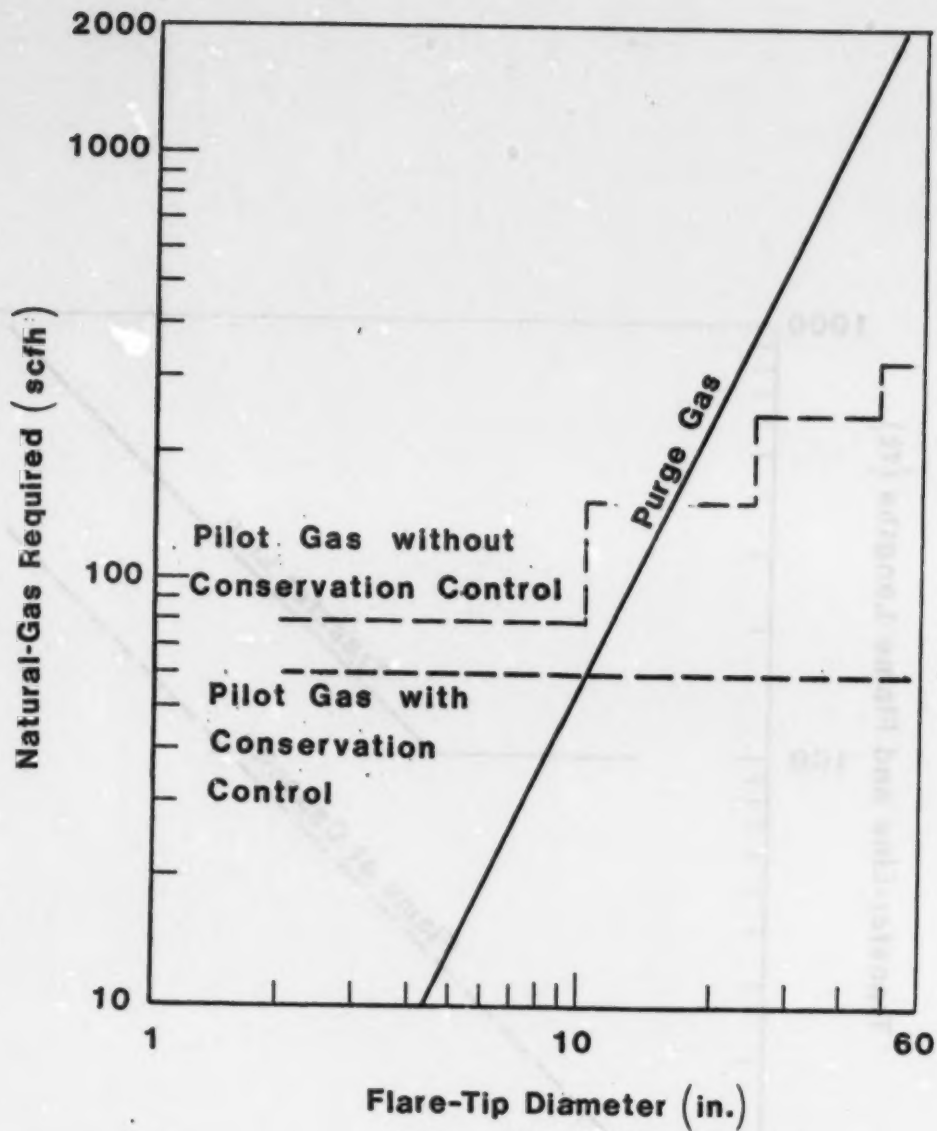


Fig. IV-4. Natural Gas Required for Elevated Flare

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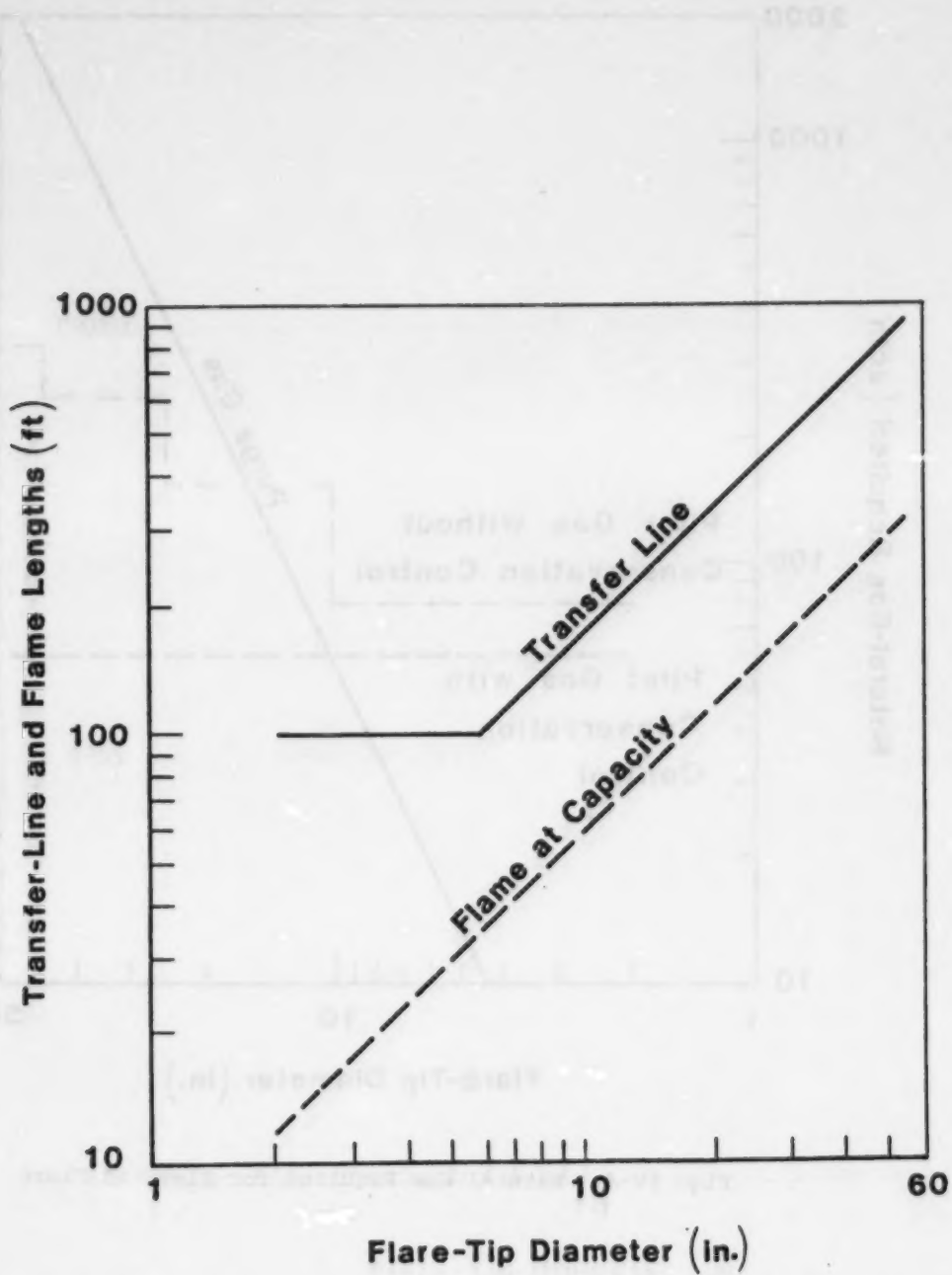


Fig. IV-5. Gas Transfer-Line and Flame Lengths for Elevated Flare

Figure IV-6 is a plot of system pressure drops for operation at smokeless capacity and at 18 in. H₂O flare-tip pressure-drop capacity. For this study it was assumed that the process emissions would have sufficient pressure to overcome the system pressure drop.

B. ENCLOSED GROUND FLARES

An enclosed ground flare consists of burner heads designed for the flare-gas properties in sufficient number to handle the maximum design flow and staged to operate based on the flare-gas flow rate. The enclosure is of sufficient cross section for the maximum heat release and is high enough to shield the illumination and radiation from the flame and to adequately disperse the heat plume. Figures IV-7, 8 and 9, based on information from a vendor for an enclosed circular type of ground flare,¹⁶ give the capacity, height, and pilot-gas requirements used in this study. It was assumed for estimating costs that the flare would be located in the area where the emission is generated and that the waste gas emission transfer-line requirement would be minimal. The location of the flare can be very site specific and the cost in some cases can be significant, especially in retrofit situations. It is estimated that 1-psig maximum gas pressure will be adequate to activate the burner head staging controls and to assure proper burner operation. It was also assumed that the process emission would have sufficient pressure to overcome the system pressure drop.

If the potential emission rate to the enclosed ground flare can overload the flare, it should have overfire protection and probably should be associated with an elevated flare or some other safety precaution to take care of the overload.

C. FUEL-GAS SOURCE

The design of a system to use an emission as a fuel-gas source is site specific. For the purposes of estimating capital and operating costs the following assumptions were made for this study: The compressor is located in the area where the emission is generated and the gas collection line required is minimal, the compression ratio is from 0 to 30 psig, the high-pressure transfer line is 500 ft, and the system capacity is the model-waste-gas flow that gives 0.5-psi pressure drop per 100 ft of the high-pressure line. Figure IV-10 is a plot of the transfer-line fuel-gas capacity.

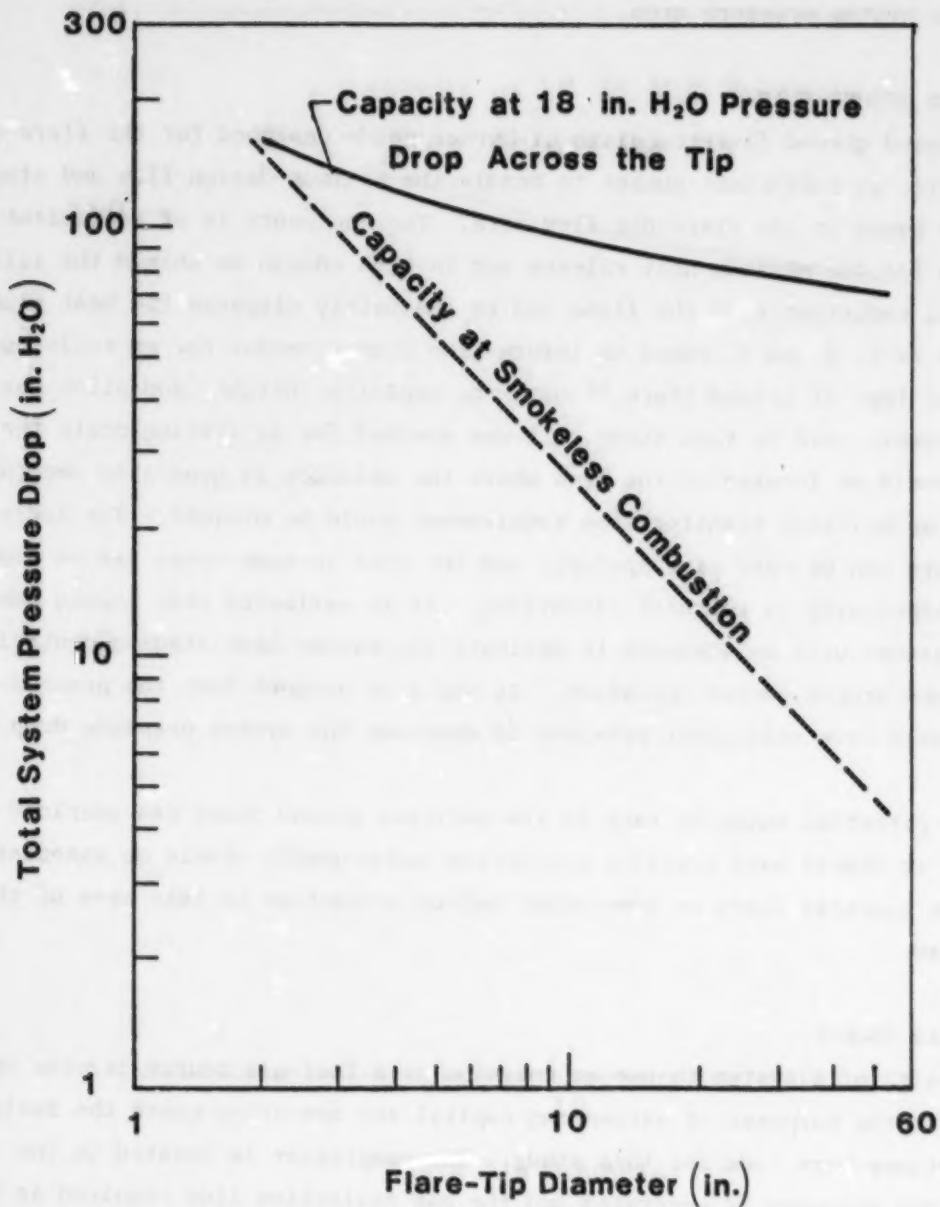


Fig. IV-6. Pressure Drop of Elevated-Flare System

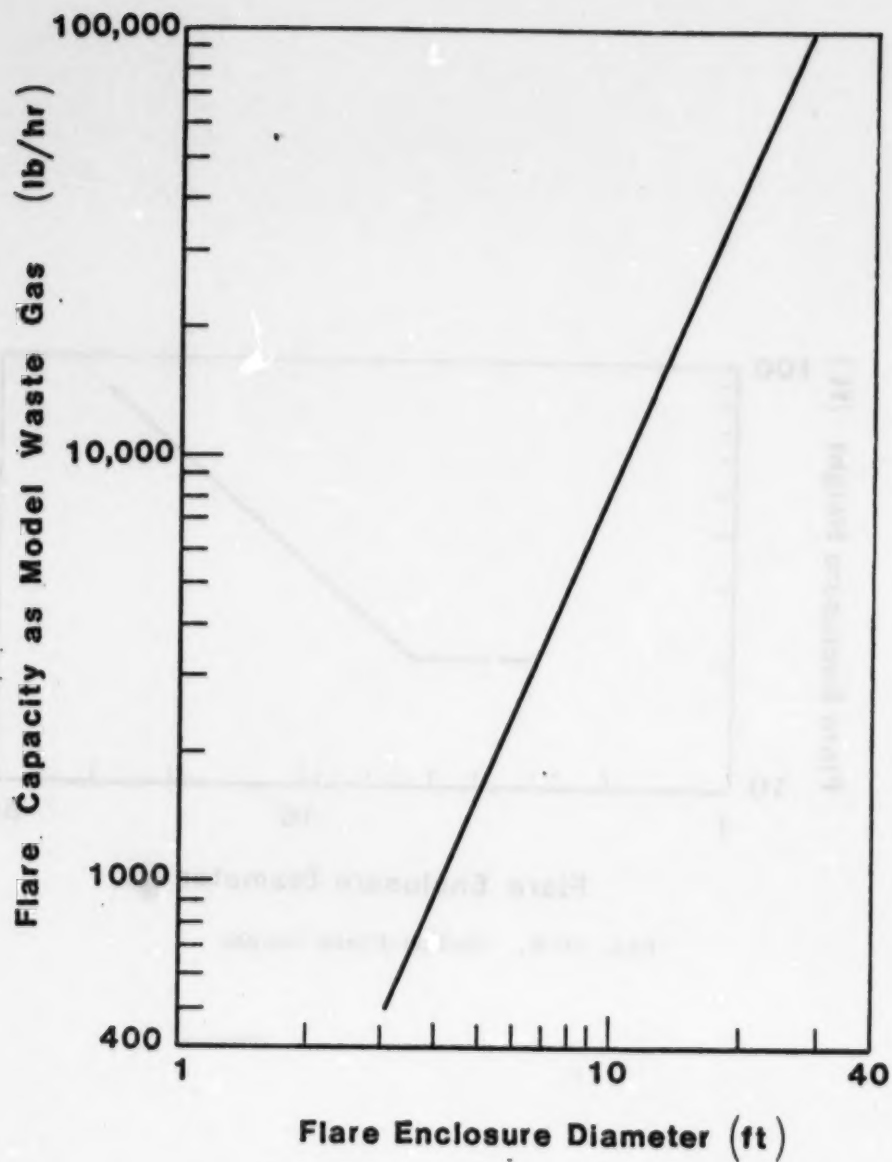


Fig. IV-7. Ground-Flare Capacity

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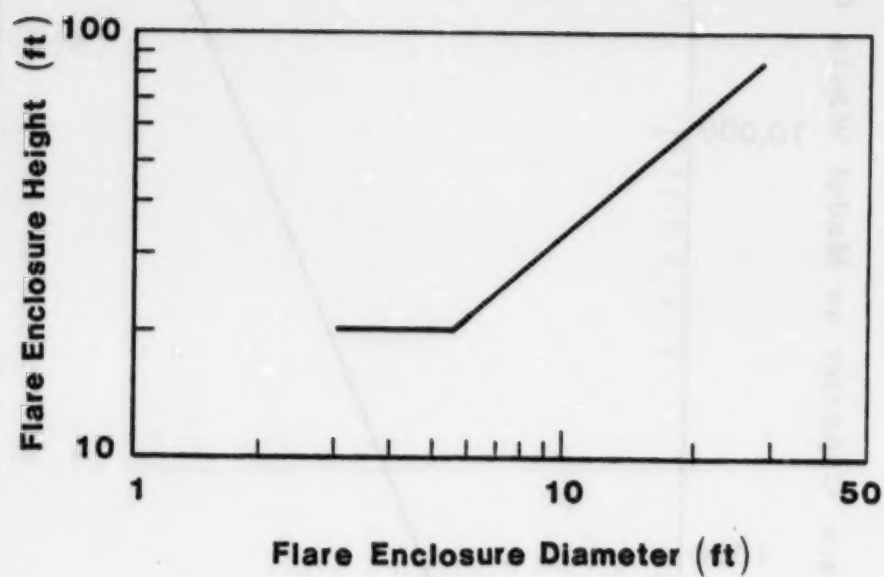


Fig. IV-8. Ground-Flare Height

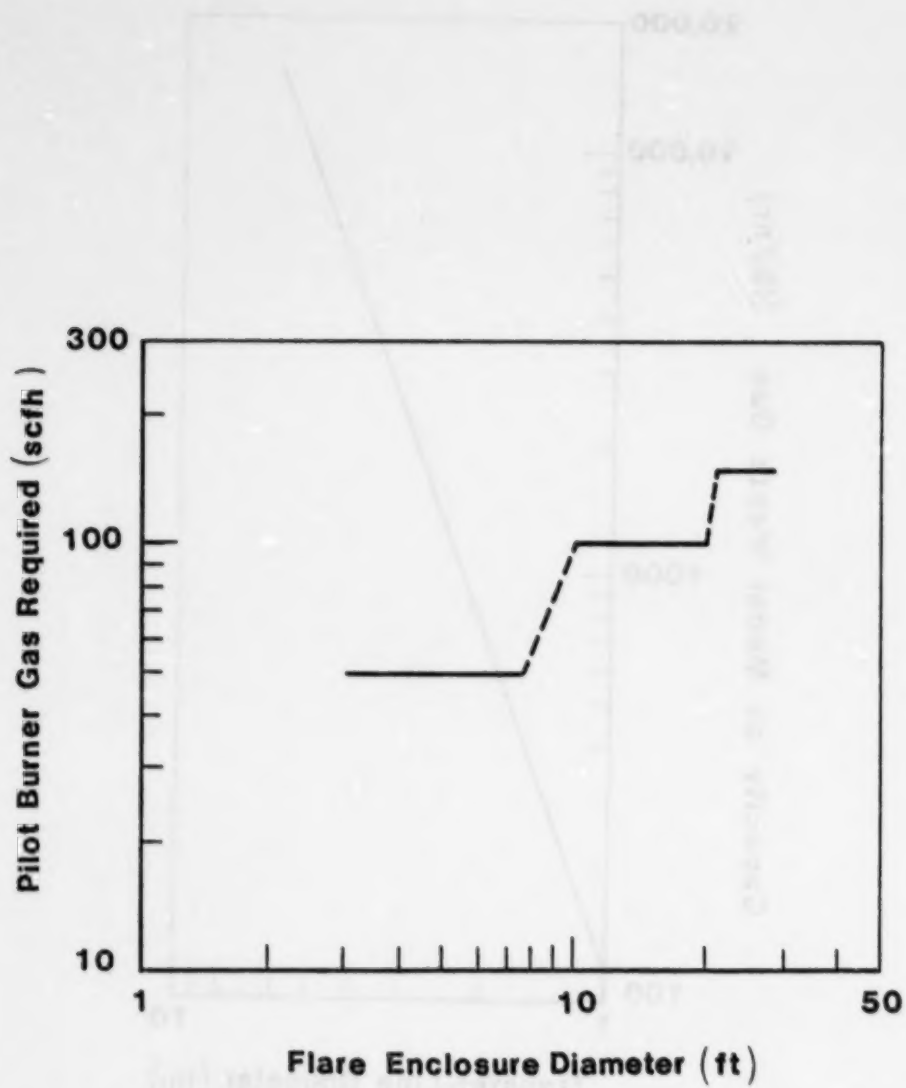


Fig. IV-9. Natural Gas Required for Ground Flare

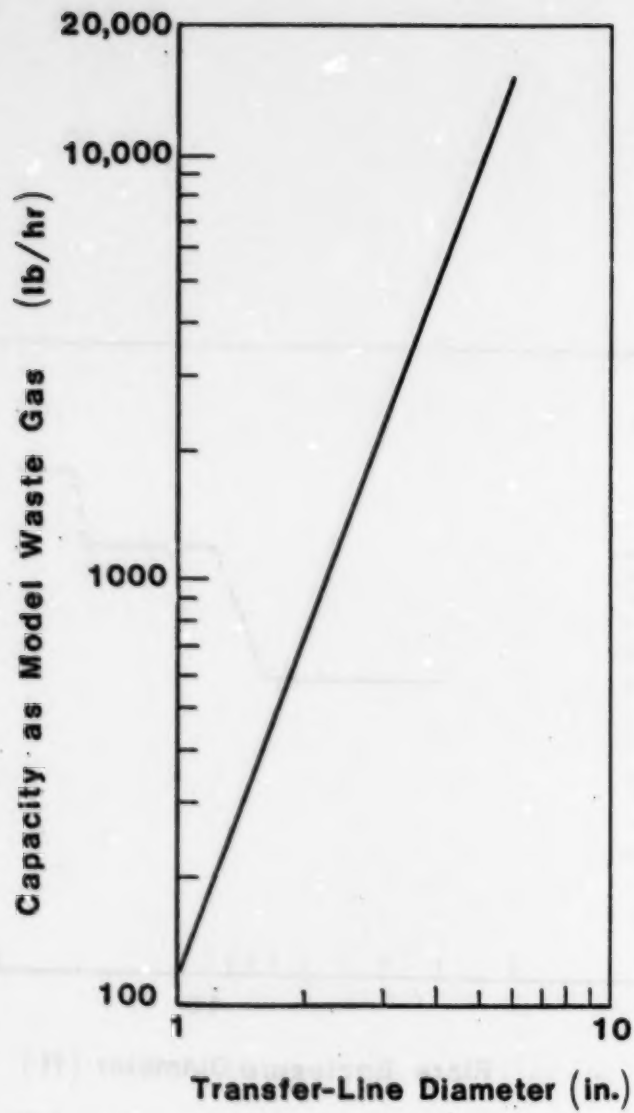


Fig. IV-10. Fuel-Gas Source Transfer-Line Capacity

If the potential emission rate to the fuel-gas source can overload the compressor or if there is not a reliable use for the emission, then an associated flare system or some other method probably should be provided to handle the emission during these periods.

V. COST AND ENERGY IMPACTS

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The capital and operating costs and the energy impacts in this section are based on the model designs discussed in Sect. IV and on a model waste gas with properties similar to those of propylene as the emission. Each control system—steam-assisted elevated flares, enclosed ground flares, and use of the emission as a fuel source—is evaluated independently. To aid in adjusting the costs presented in this section to other situations that may vary considerably from the model conditions assumed, the installed capital costs of each system are presented based on the physical sizes shown in Appendix B.

The estimated capital costs represent the total investment, including all indirect costs such as engineering and contractors' fees and overheads that will be required for purchase and installation of all equipment and material to provide a facility as described. These are battery-limit costs and do not include the provisions for bringing utilities, services, or roads to the site, the backup facilities, the land, the research and development required, or the process piping and instrumentation interconnections that may be required within the process generating the waste gas. These costs are based on a new-plant installation; no retrofit cost considerations are included. Those costs are usually higher than the cost for a new-site installation for the same system and include, for example, demolition, crowded construction working conditions, scheduling construction activities with production activities, and longer interconnecting piping. These factors are so site specific that no attempt has been made to provide costs. For specific retrofit cases, rough costs can be obtained by using the new-site data and adding as required for a defined specific retrofit situation.

The method used to develop these estimated capital costs was based on preliminary vendor quotes for the purchase and installation of major equipment items or from such sources as Richardson Engineering Co. data and factoring up to total installed costs based on the factors in Table V-1. The expected accuracy of the total installed cost with this degree of engineering detail using these methods is $\pm 30\%$. These methods of obtaining estimated total installed capital costs are suitable for a cost study or for screening estimates. The bases used in calculating annual operating costs are given in Table V-2.

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Table V-1. Factors Used for Estimating Total Installed Costs

A = Major equipment purchase cost plus 0.1 to 0.35 allowance

Installation costs

Foundations	0.06A + \$100 X number of pumps
Structures	0.15A (no structures) to 0.30A (multideck structures)
Equipment erection	0.15A to 0.30A (depending on complexity)
Piping	0.40A (package units) to 1.10A (rat's nest)
Insulation	0.06A or 0.15 X piping (normal) to 0.30 X piping (bulk hot or cold)
Paint	0.05A
Fire protection	0.01A to 0.06A (depending on requirements)
Instruments	0.10A to 0.30A or 0.01A to 0.25A + \$50,000 to \$300,000 for process control computer
Electrical	0.15A or 0.05A + \$500 per motor

B = Base cost	A + sum of installation costs
Sales tax	0.25A + 0.025B
Freight	0.16A
Contractor's fees	0.30 (B-A)
C = Total contract	B + taxes, freight, and fees
Engineering ^a	0.01C to 0.20C
Contingencies ^b	0.15C

D = Process unit installed cost C + engineering + contingencies

E = Total subestimates

Sum of semidetalled subestimates (buildings, site development, cooling towers, etc.); each subestimate should include taxes, freight, fees, engineering, and contingency, and should be escalated to date of expenditure for that cost component; engineering costs, contingencies, and escalation factors for these subestimates will vary according to the type of job involved

F = Total project cost D + E

^aIncludes cost from capital project teams, process engineering, engineering, purchasing, and other support groups.

^bContingency should not be applied to any cost component that has been committed by either purchase order or contract.

Table V-2. Annual Cost Parameters

Operating factor	8760 hr/yr ^a
Operating labor	Negligible
Fixed costs	
Maintenance labor plus materials, 6%	} 29% of installed capital cost
Capital recovery, 18% ^b	
Taxes, insurances, administration charges, 5%	
Utilities	
Electric power	\$0.03/kWh
Steam	\$2.50/thousand lb
Natural gas	\$2.00/thousand ft ³
Heat recovery credits (equivalent to natural gas)	\$2.00/million Btu

^aControl devices will usually be available for operation on the same cycle as the process. Process downtime is normally expected to range from 5 to 15%. From the standpoint of cost-effectiveness calculations the error introduced by assuming continuous availability is minor. The percent of capacity operation that the control device is utilized will be discussed as a variable in this section.

^bBased on 10-year life and 12% interest.

A. STEAM-ASSISTED ELEVATED FLARES

The installed capital costs, annual operating costs, cost-effectiveness, and energy-effectiveness curves for the model systems based on the model waste gas as the VOC emissions are shown in Figs. V-1 to V-4 respectively. For installed capital costs based on flare size see Appendix B.

The annual operating cost will vary with the capacity of operation because steam assistance is used only during the time that flaring takes place. Normally an elevated flare will operate for only a very small percentage of the time; even so, steam costs can be an important factor. In this study it was assumed that 150-psig steam is available to the flare on demand at a nominal cost of \$2.50 per thousand pounds.

The cost effectiveness and energy effectiveness of elevated flares are sensitive to the capacity of operation in the low ranges of operation that are normal. One-tenth percent of annual capacity operation is equal to about 9 hr per year of flaring at smokeless capacity, or 18 hr at half capacity. Natural gas used for purge gas and for the pilots and steam used for injection at the flare tip are the sources of energy consumption. It was assumed that modern flare-tip seal designs would be used to minimize the need for purge gas and that instrumentation would be used to reduce pilot-gas consumption. The deviation in cost and energy effectiveness shown on Fig. V-3 and V-4 at the larger capacities results from a higher ratio of steam to model waste gas being used for flares larger than 24-inch tip diameter and from the fact that purge gas consumption increases as the cross section of the flare while the smokeless capacity proportionately decreases because of steam availability. This effect is more pronounced for the lower percent of annual capacity operation and for the energy effectiveness ratio.

The cost and energy effectiveness presented here is based on the flare being used for pollution control. This must be evaluated with caution in many actual situations since elevated flares are frequently designed and installed as safety devices.

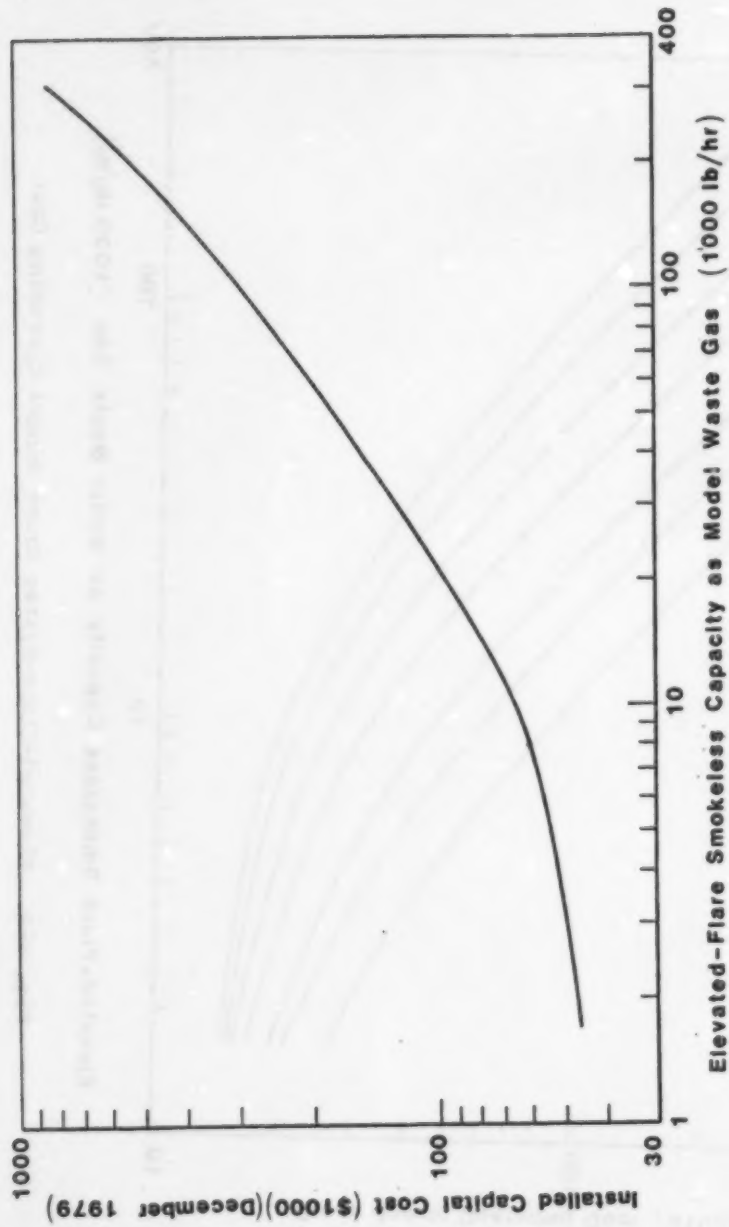


Fig. V-1. Elevated-Flare-System Installed Capital Cost

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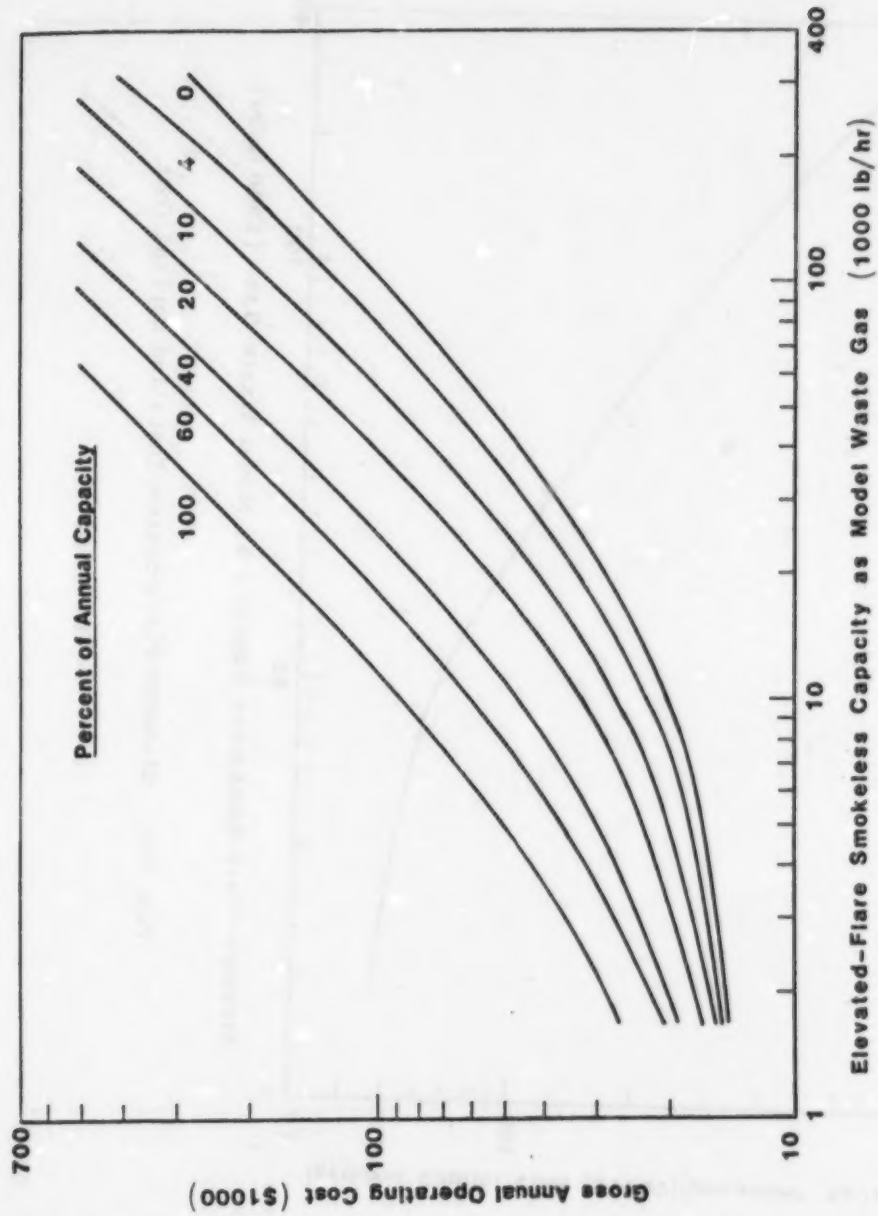


Fig. V-2. Elevated-Flare-System Gross Annual Operating Cost

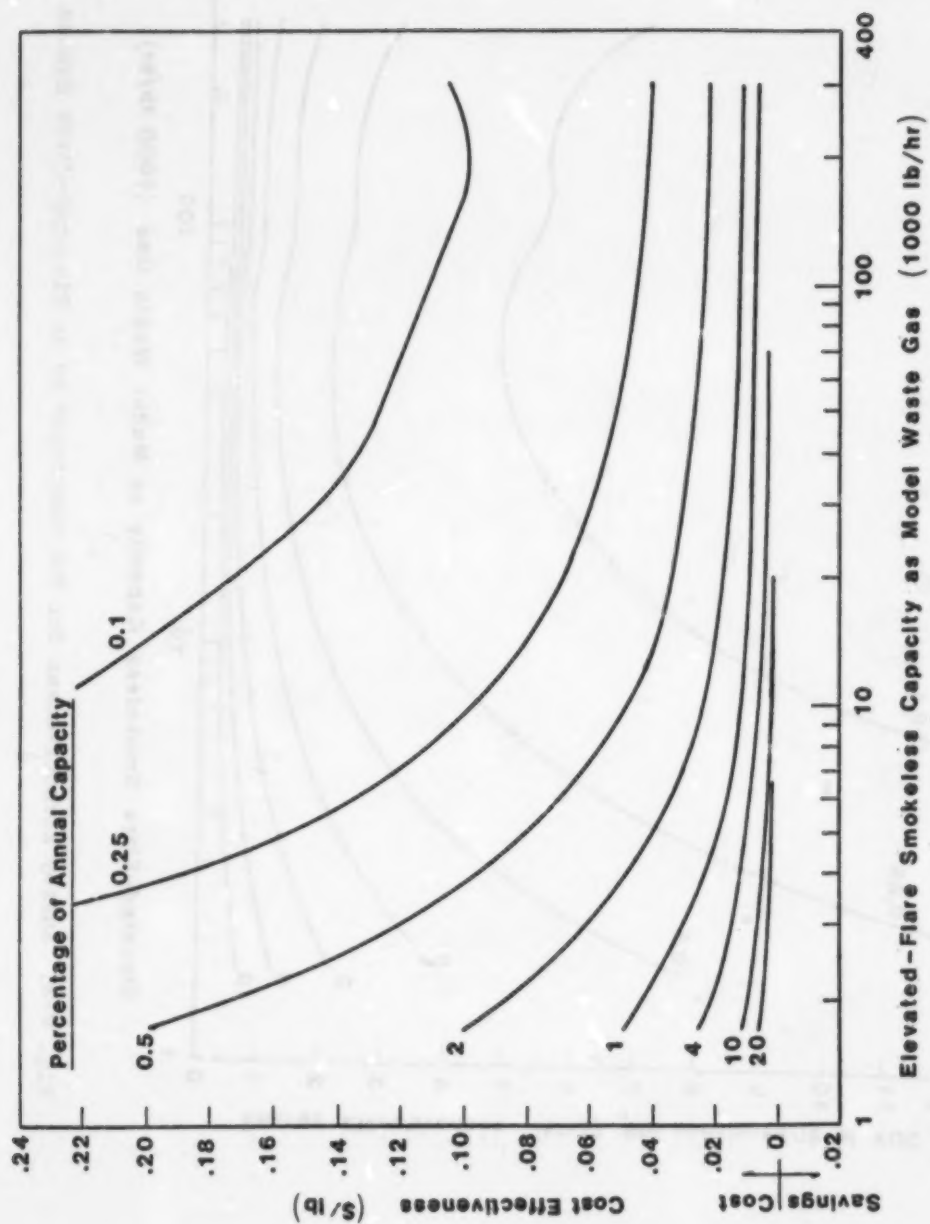


Fig. V-3. Cost Effectiveness of VOC Destroyed by an Elevated-Flare System

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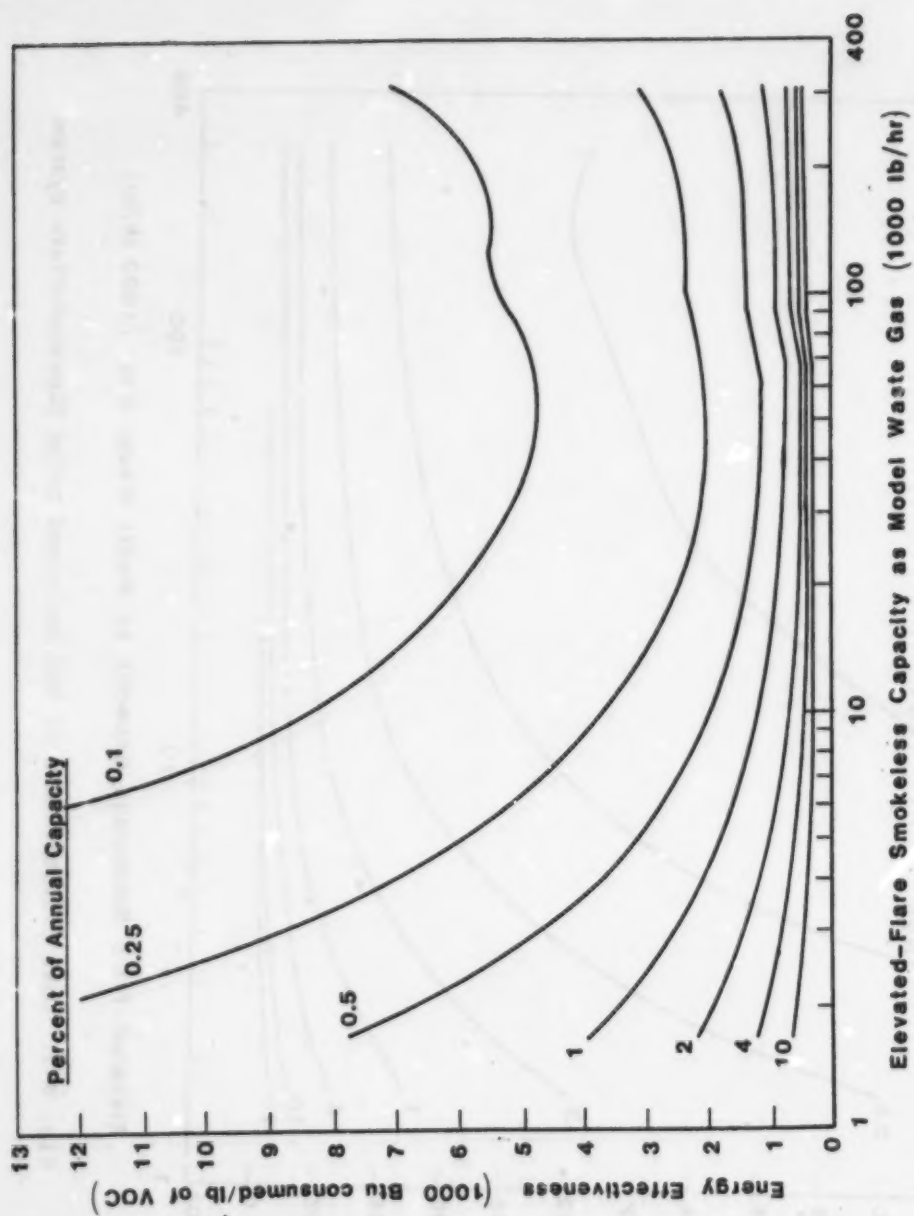


Fig. V-4. Energy Effectiveness for VOC Destroyed by an Elevated-Flare System

B. ENCLOSED GROUND FLARES

The installed capital costs, annual operating costs, cost-effectiveness, and energy-effectiveness curves for the model systems based on the model waste gas as the VOC emission are plotted in Figs. V-5 to V-8 respectively. For installed capital costs based on flare size see Appendix B.

The cost effectiveness and energy effectiveness of ground flares are sensitive to the capacity of operation in the low ranges of operation that are probable. A 0.25% of annual capacity operation is equal to about 22 hr per year of flaring at capacity, or 44 hr at half capacity. The discontinuities in the energy-effectiveness curves reflect the need for additional pilot burners as the flare increases in size.

C. FUEL-GAS SOURCE

The installed capital cost, annual operating costs, and cost-effectiveness curves for the model systems based on the model waste gas are shown in Figs. V-9 to V-11 respectively. For installed capital costs based on the fuel-gas-line size see Appendix B.

The costs associated with the use of VOC emission as a fuel gas are very site specific. The costs presented here are valid only for the models described but are descriptions of what may be expected. There is a cost-effectiveness savings, except for low-capacity operation, when the VOC emission is credited at a fuel-gas value equivalent to \$0.0392 per pound of model waste gas (\$2.00/million Btu). Use of VOC as a fuel may be an attractive disposal method but only if the emission meets the requirements of being satisfactory for use as a fuel, is reasonably consistent in generation, and there is an adequate use for it at a reasonable distance.

Using a VOC emission as a fuel can be very energy effective. For the model system about 40 Btu of electrical energy is consumed per pound of model waste gas compressed, whereas the net heat of combustion for the model waste gas is 19,600 Btu/lb.

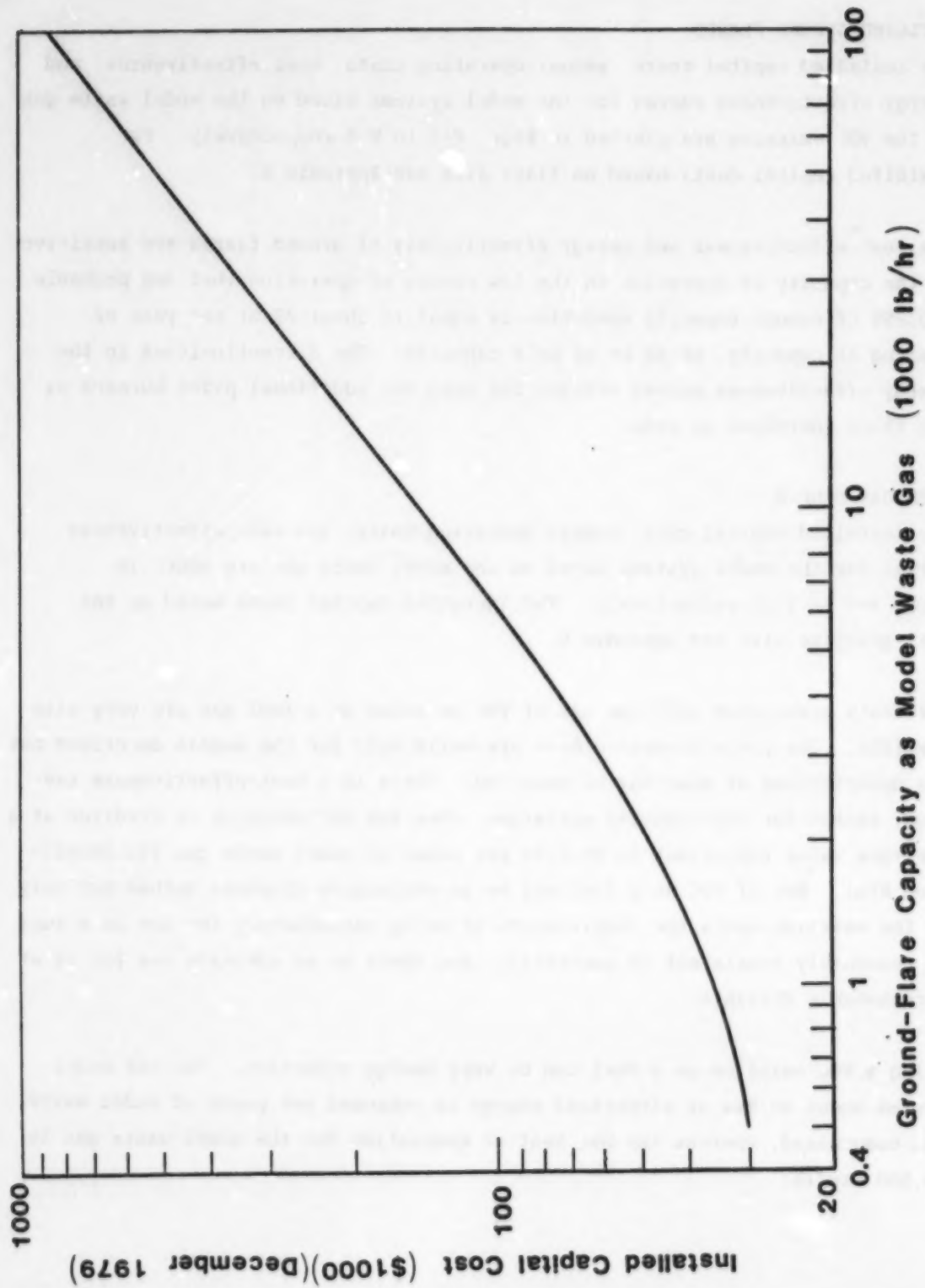


Fig. V-5. Ground-Flare-System Installed Capital Cost

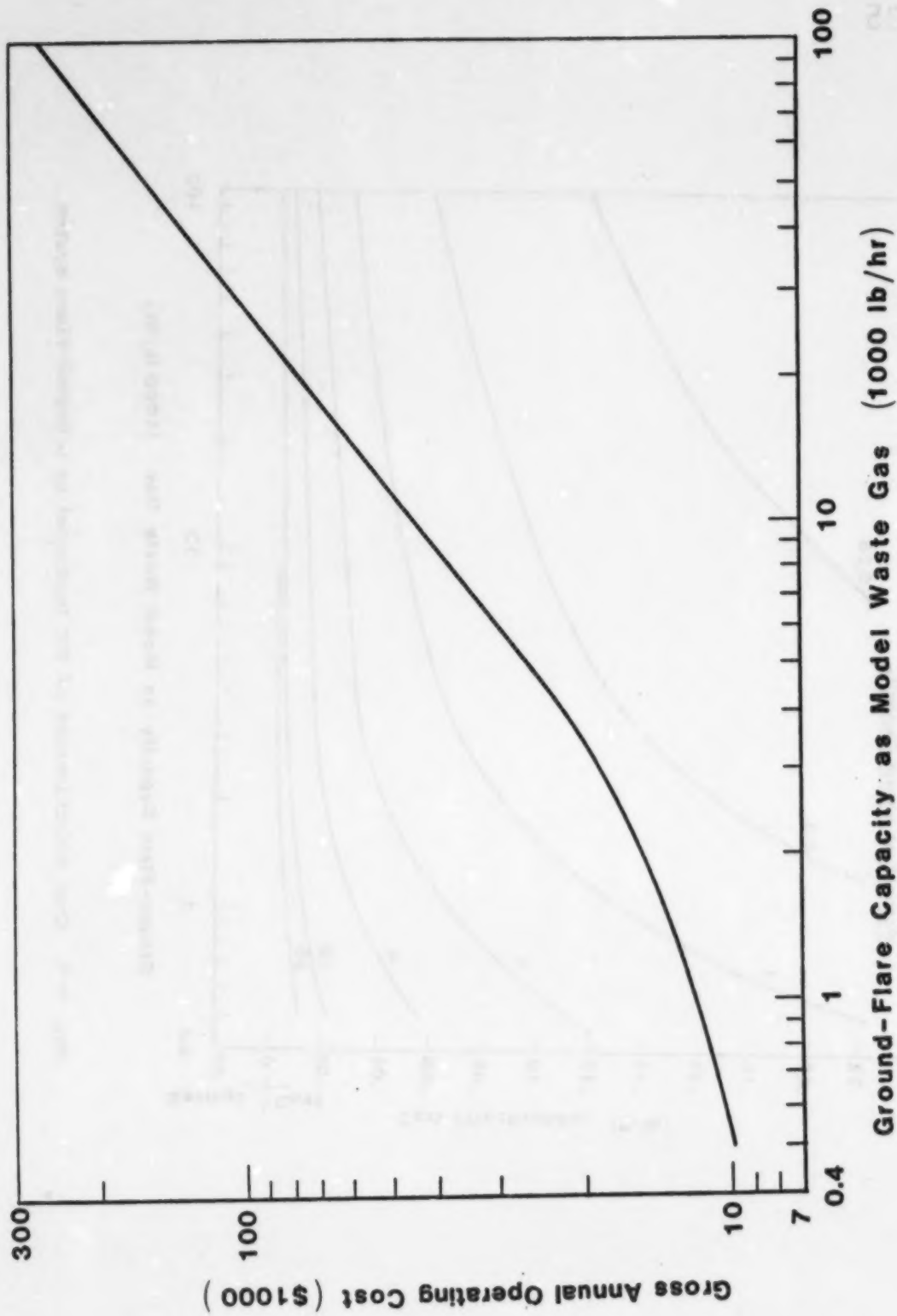


Fig. V-6. Ground-Flare-System Gross Annual Operating Cost

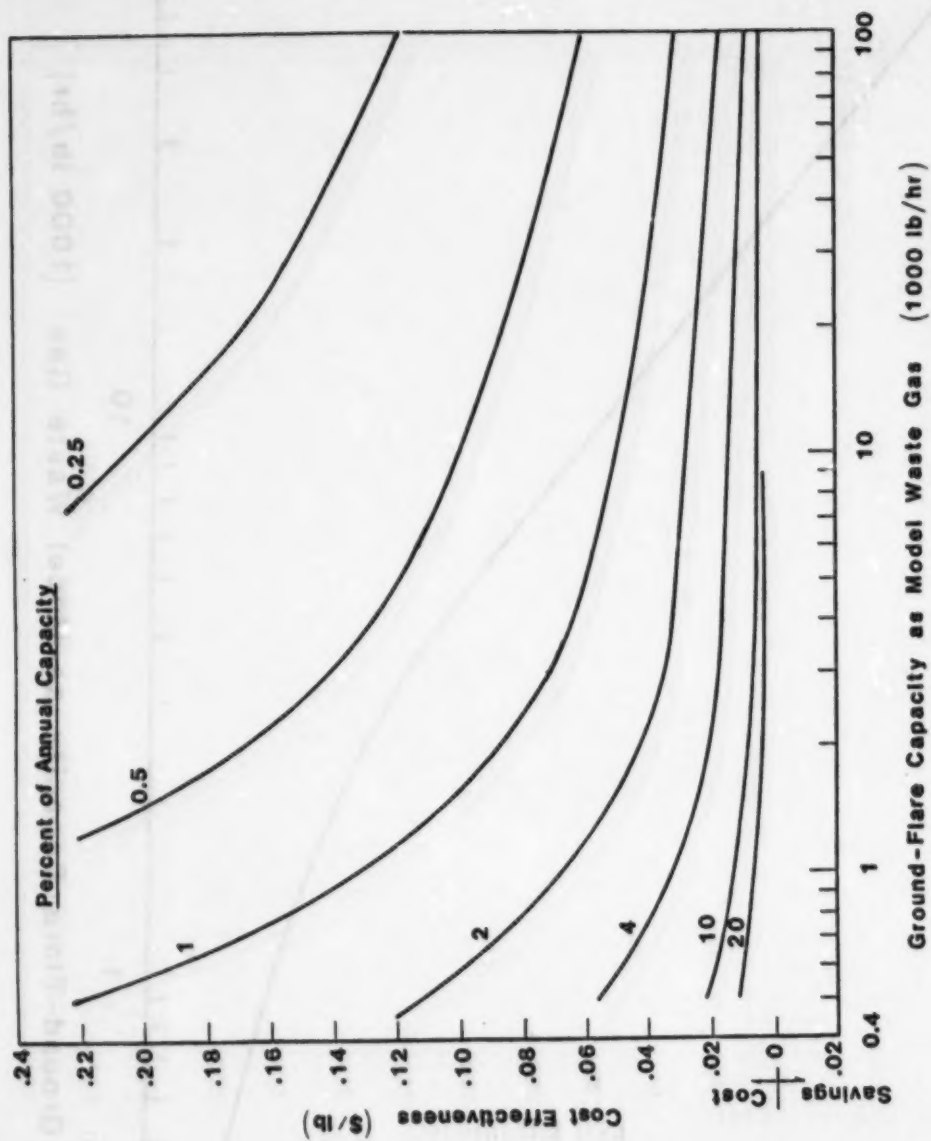


Fig. V-7. Cost Effectiveness of VOC Destroyed by a Ground-Flare System

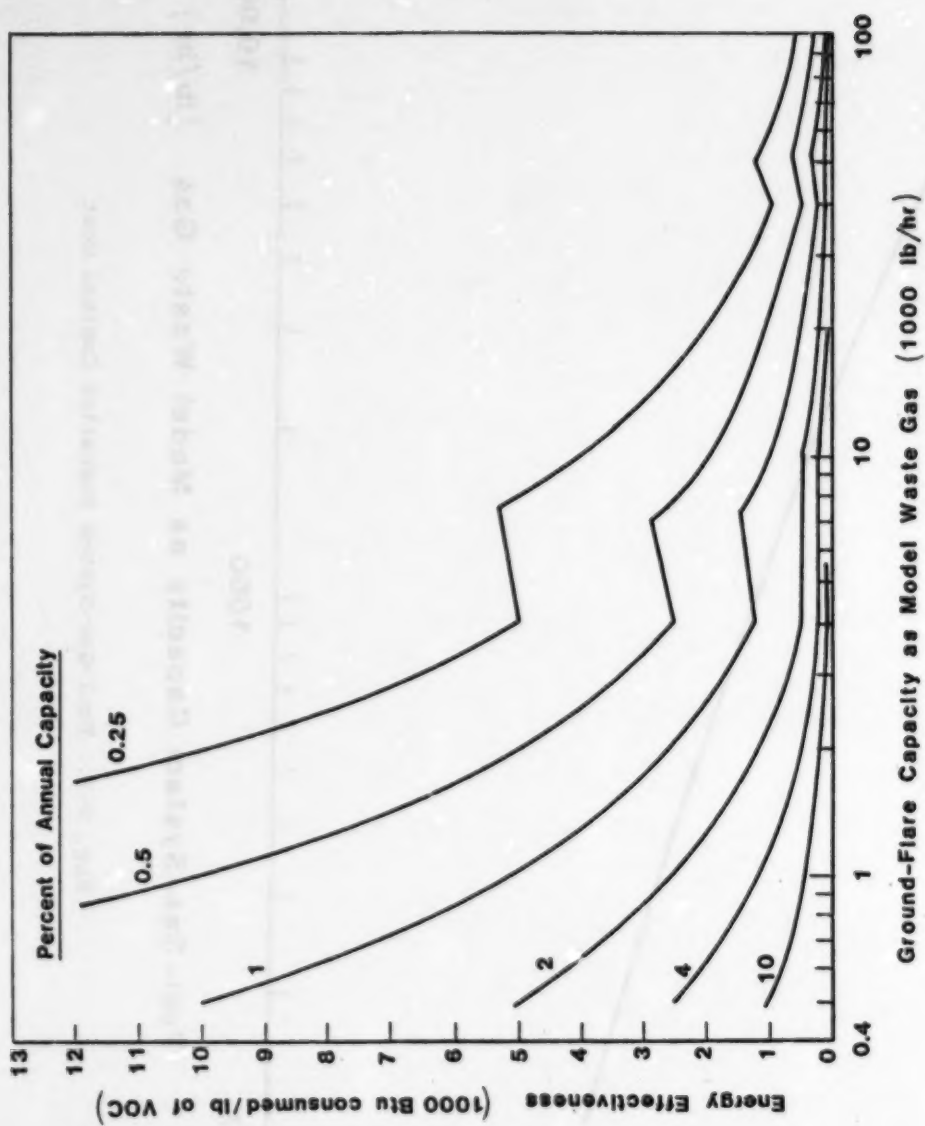


Fig. V-8. Energy Effectiveness for VOC Destroyed by a Ground-Flare System

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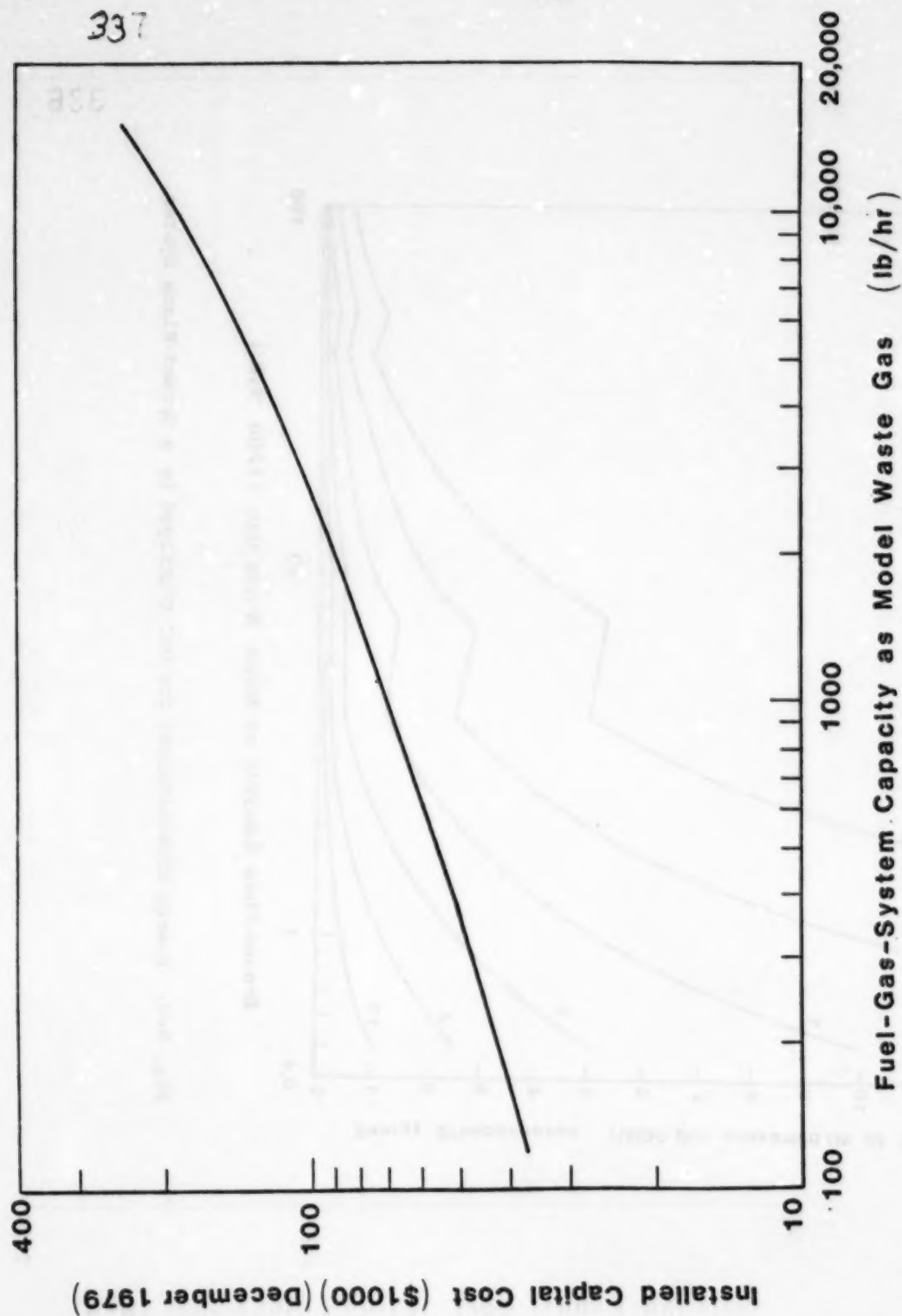


Fig. V-9. Fuel-Gas-System Installed Capital Cost

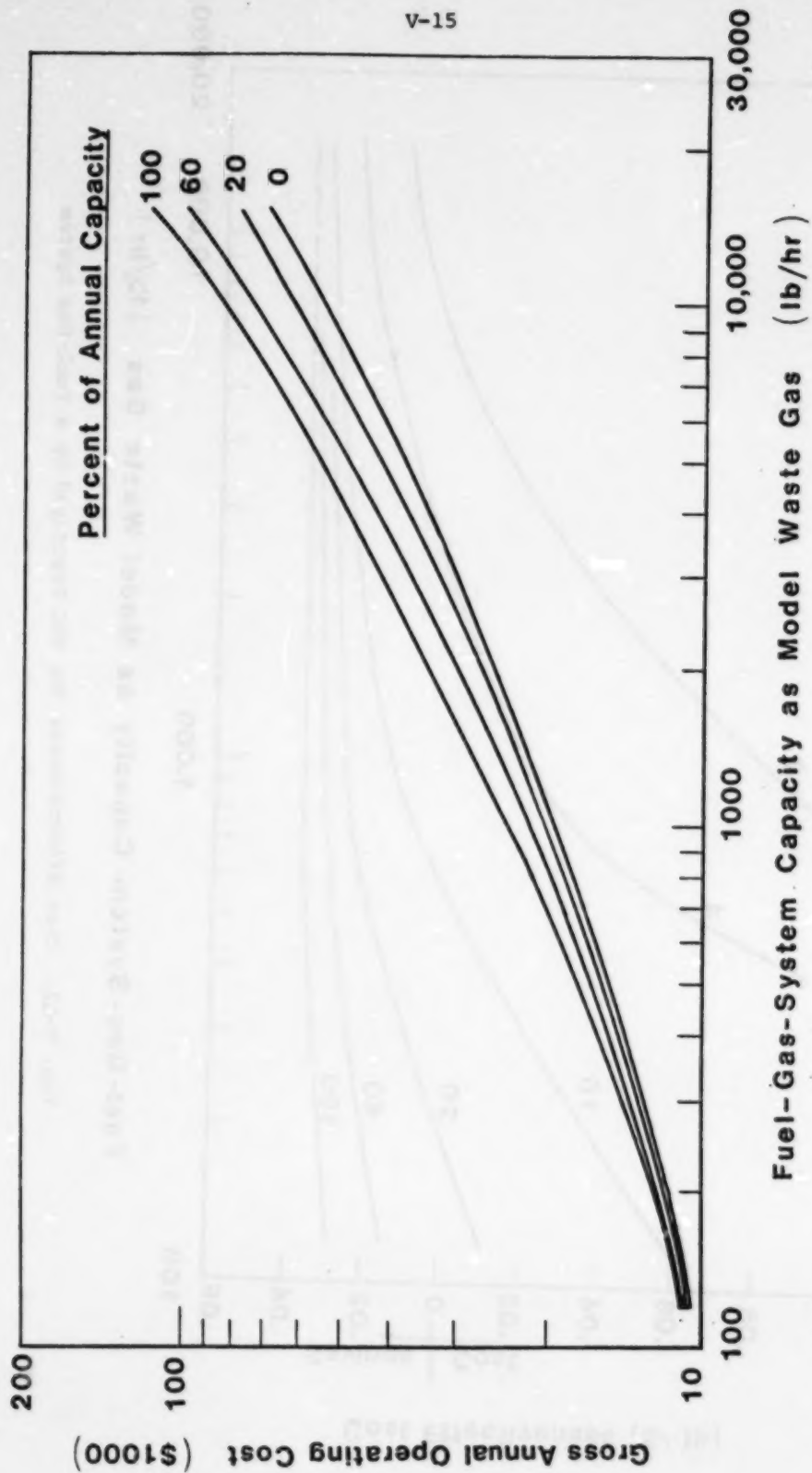


Fig. V-10. Fuel-Gas-System Gross Annual Operating Cost

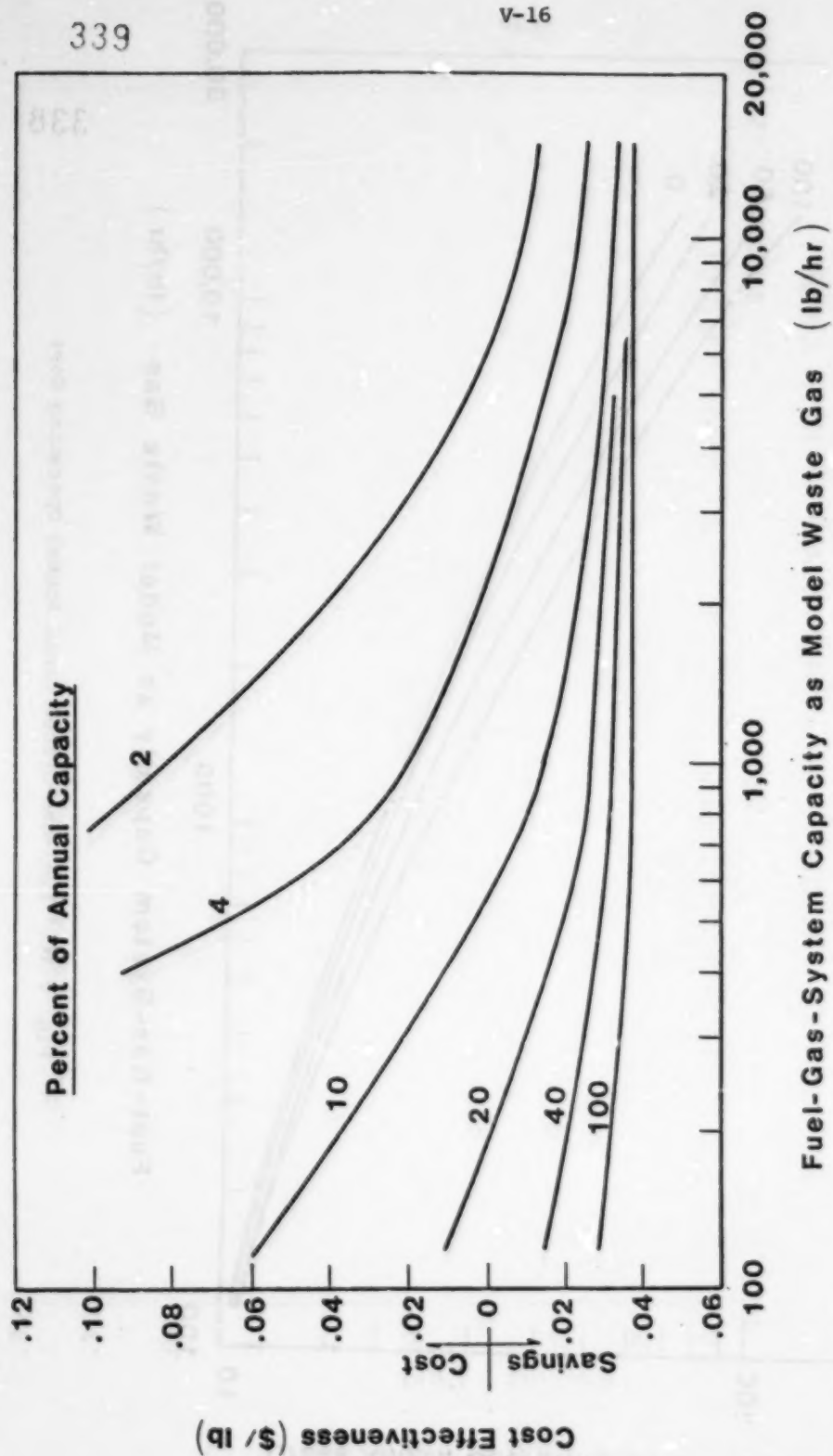


Fig. V-11. Cost Effectiveness for VOC Destroyed by a Fuel-Gas System

VI. SUMMARY AND CONCLUSIONS

The emission control methods of burning VOC emissions in an elevated flare, in a ground flare, or using it as a fuel have been evaluated in this study. Flares frequently are installed primarily for the purpose of safety. Each control method has its advantages and its limitations and could, for appropriate emission sources, be installed in an integrated system. Such integrated systems would be very site specific and are beyond the scope of this study.

A. STEAM-ASSISTED ELEVATED FLARES

1. Steam-assisted elevated flares can be designed for large capacities and can take overloads, but the smokeless capacity is usually limited to the amount of steam available.
2. Operation can be highly transitory.
3. The data on VOC destruction efficiency are limited. This study for cost-effectiveness presentation purposes only is based on an efficiency of 99% for flare tips under 12-in. diameter and 98% for those over 12-in. diameter. The EPA is conducting further testing.
4. An elevated flare must be ready for use at any time that there is potential for emissions, but it may be actually flaring for a very small percent of the time. Under these circumstances the energy consumption per quantity of VOC destroyed could be high.

B. ENCLOSED GROUND FLARES

1. Enclosed ground flares can be designed for a wide capacity range but may need overload protection if potential for overloading exists.
2. Operation can be highly transitory, but 100 to 0% of capacity turndown exists.
3. VOC destruction efficiency data are not available; a 99% destruction was used to calculate cost effectiveness.
4. When a ground flare is used to burn emissions for only a small percent of its available time, the energy consumption per quantity of VOC destroyed can be high. The energy use can be comparable to that of an elevated flare that is designed and operated to minimize natural-gas consumption.

C. FUEL-GAS SOURCE

1. Using VOC emission as a fuel can give a cost-effectiveness savings provided that the emission is of fuel quality, is produced in a reasonably consistent volume, and there is an adequate use for it.
2. The VOC destruction efficiency used for cost-effectiveness calculations was >99.9%. According to emission factors given in AP-42 for natural gas and LPG when burned under proper operating conditions, this efficiency is attained.

VII. REFERENCES*

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2. M. G. Klett, Lockheed Missiles and Space Company, Report on a Trip to the John Zink Company to Discuss EPA Tank 3, Flare Systems Study, Feb. 19, 1974.
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10. Personal communication Feb. 25, 1980, between John F. Straitz, III, National Air Oil Burner Company, Inc., and V. Kalcevic, IT Envirosience.
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14. Robert D. Reed, "Design and Operation of Flare Systems," Chemical Engineering Progress (June 1968).
15. John F. Straitz, III, National Air Oil Burner Company, Inc., letter dated Oct. 21, 1980, to EPA with comments on draft report.
16. John F. Straitz, III, Combustion Unlimited, Inc., letter dated June 5, 1980, to V. Kalcevic, IT Envirosience, Inc.

*When a reference number is used at the end of a paragraph or on a heading, it usually refers to the entire paragraph or material under the heading. When, however, an additional reference is required for only a certain portion of the paragraph or captioned material, the earlier reference number may not apply to that particular portion.

APPENDIX A**ELEVATED-FLARE DESIGN EQUATIONS**

A. FLARE CAPACITIES AND PRESSURE DROPS

Flare capacities and pressure drops used in this report are based on the following relationship:^{9*}

$$D = \left[\frac{2.72 \times 10^{-3} (F) \sqrt{\frac{T + 460}{MW}}}{\sqrt{\Delta p}} \right]^{\frac{1}{2}},$$

where

D = flare-tip diameter, in.,

F = flare gas flow rate, lb/hr,

T = temperature, °F (assumed to be 60°F),

MW = molecular weight (propylene, 42),

Δp = flare-tip pressure drop, in. H₂O.

B. FLARE HEIGHTS

The flare heights used in this report are based on the following relationship:¹²

$$H = \sqrt{\frac{F \times LHV \times \epsilon}{12.56 I}} - 3.33 D \sqrt{\frac{\Delta p}{55}} \cos \theta,$$

where

$$\theta = \tan^{-1} \frac{1.47 V_w}{550 \sqrt{\frac{\Delta p}{55}}},$$

in which V_w = wind velocity, mph (assumed to be 60),

H = flare height, ft,

F = flare gas flow rate, lb/hr,

LHV = flare gas lower heating value, Btu/lb (propylene, 19,600).

ϵ = flame emissivity (propylene, 0.13),

I = flame radiation intensity, Btu/(hr)(ft²) (assumed to be 1200),

D = flare-tip diameter, in.,

Δp = flare-tip pressure drop, in. H₂O.

*See Sect. VII for references.

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C. FLAME LENGTHS

The flame lengths used in this report to determine the gas transfer-line lengths are based on the following relationship:³

$$L_f = 10 \times D \times \left(\frac{\Delta p}{55} \right)^{1/2}$$

where

L_f = flame length, ft,

D = flare-tip diameter, in.,

Δp = flare-tip pressure drop, in. H_2O (assumed to be 18).

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APPENDIX B

INSTALLED CAPITAL COSTS

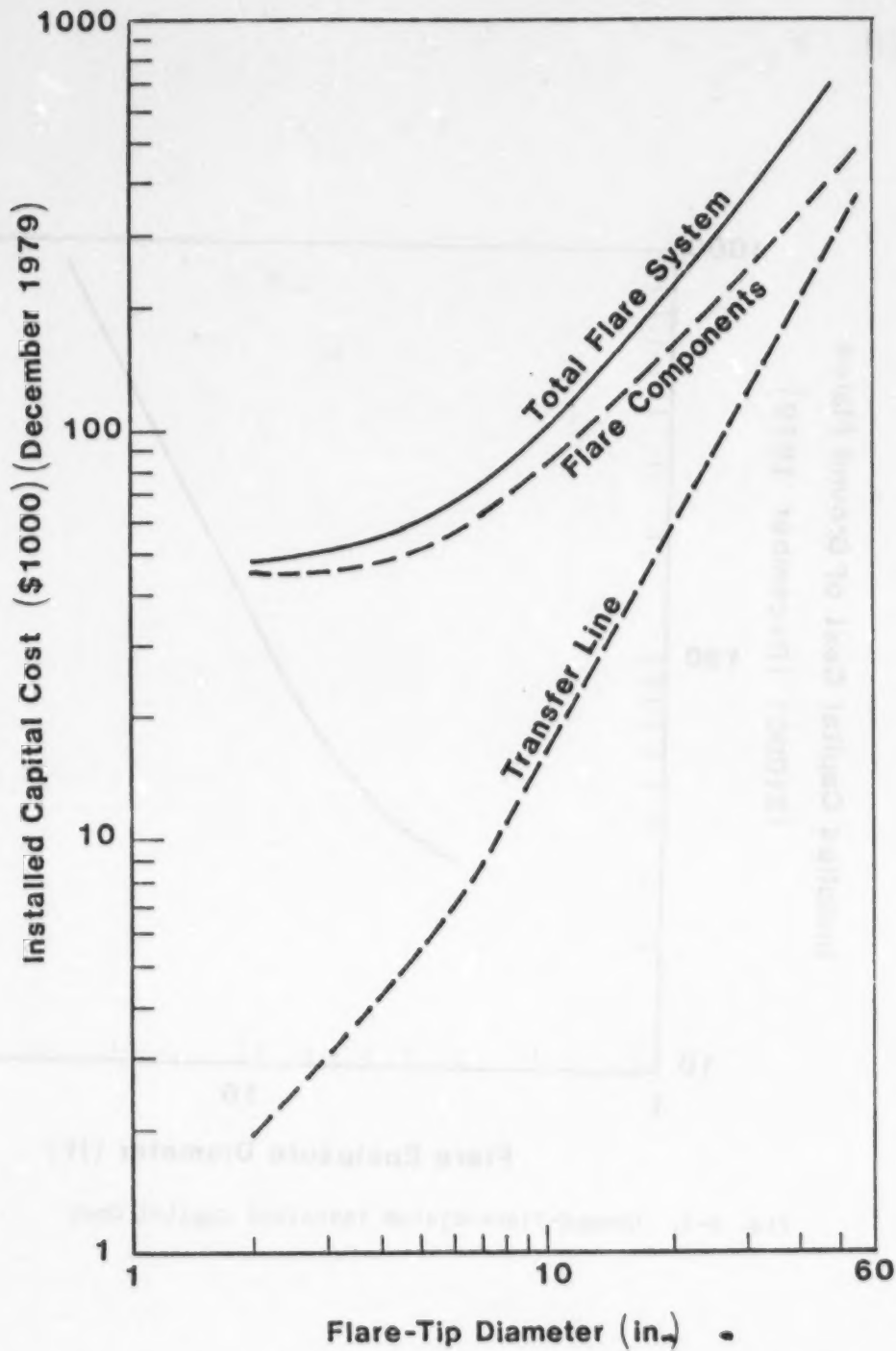


Fig. B-1. Elevated-Flare-System Installed Capital Cost

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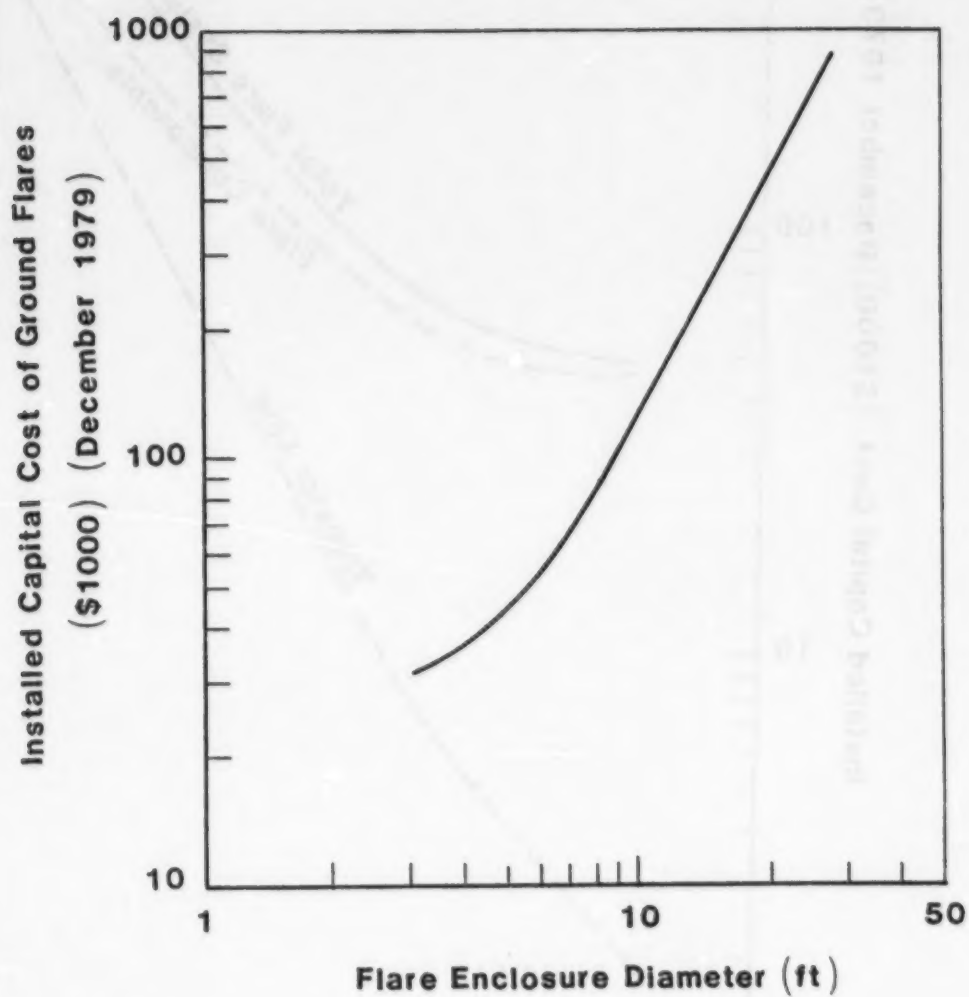


Fig. B-2. Ground-Flare-System Installed Capital Cost

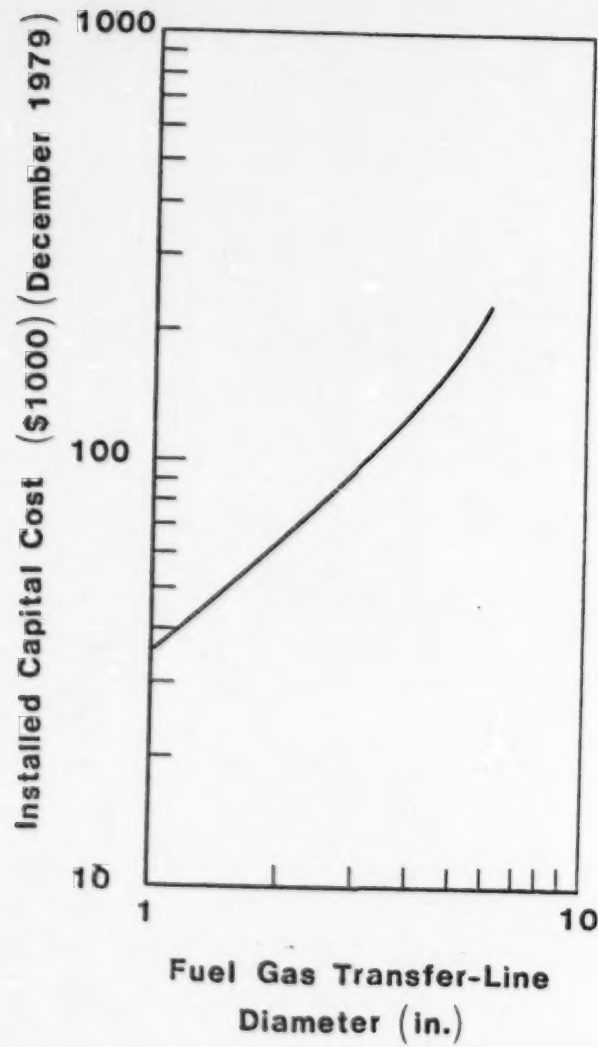


Fig. B-3. Fuel-Gas-System Installed Capital Cost

APPENDIX C

SAMPLE CALCULATIONS

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COST AND ENERGY SAMPLE CALCULATIONS

This report is based on using a model waste gas with properties similar to those of propylene as the VOC emissions. The curves presented can be used in most cases for other VOC emissions with results probably within the accuracy of the curves developed. For emissions that have a character significantly different from the model ones it may be advisable to review the design bases discussed in Sect. IV and then use the installed capital costs given in Appendix B.

A. STEAM-ASSISTED ELEVATED FLARES

This example is based on a 23,000-lb/hr maximum model-waste-gas VOC emission rate.

1. Installed Capital Cost Versus Flare Capacity (Fig. V-1)

- a. From Fig. IV-1, a 10-in. tip diameter flare is required.
- b. From Fig. B-1, the installed capital cost for a 10-in. flare is \$105,000.

2. Gross Annual Operating Cost (Fig. V-2)

- a. From Table V-2 the fixed cost, including capital recovery, is 29% of the installed capital cost:

$$105,000 \times 0.29 = \$30,450/\text{yr.}$$

- b. From Fig. IV-4 the natural gas used for the pilots is 60 scfh and for purging it is 60 scfh. From Table V-2 the cost for gas is \$2.00 per thousand ft³:

$$(60 + 60) \times 8760 \times \frac{2.00}{1000} = \$2100/\text{yr.}$$

- c. From Sect. IV-A-1 of the report 0.3 lb of steam is required per pound of VOC; from Table V-2 the cost for steam is \$2.50/thousand lb:

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$$0.3 \times 23,000 \times 8760 \times \frac{2.50}{1000} \times \text{fraction of annual capacity operation}$$

$$= 151,000 \times \text{fraction of annual capacity operation.}$$

d. Annual cost summary

	Cost per Capacity Operation of		
	100%	1%	0%
Fixed	\$ 30,450	\$30,450	\$30,450
Gas	2,100	2,100	2,100
Steam	<u>151,100</u>	<u>1,510</u>	<u> </u>
Total	\$183,650	\$34,060	\$32,550

3. Cost Effectiveness (Fig. V-3)

Cost effectiveness is the gross annual operating cost (A-2-d above) divided by the annual waste-gas VOC destroyed at 99% efficiency.

a. Annual VOC destroyed = $23,000 \times 8760 \times 0.99 \times \text{annual fraction of capacity operation}$.

b. Example at 1% of annual capacity operation is

$$\frac{34,060}{23,000 \times 8760 \times 0.99 \times 0.01} = \$0.017/\text{lb of VOC destroyed.}$$

4. Energy Effectiveness (Fig. V-4)

Energy effectiveness is the energy consumed in the gas pilots, the purging gas, and the steam-assist gas divided by the annual waste-gas VOC destroyed at 99%, or 98% efficiency.

a. From A-2-b, natural gas consumed is 120 scfh, at 1000 Btu/ft³ = 120,000 Btu/hr.

b. From A-3-c, steam consumed is 0.3 lb/lb of waste-gas VOC, at 1000 Btu/lb of steam; this results in 300 Btu/lb of waste-gas VOC.

- c. Example at 1% of annual capacity operation:

$$\left(\frac{120,000 \times 8760}{23,000 \times 8760 \times 0.01} + 300 \right) \times \frac{1}{0.99} = 830 \text{ Btu/lb of Voc destroyed.}$$

B. ENCLOSED GROUND FLARES

This example is based on 20,000-lb/hr maximum waste-gas VOC emission rate.

1. Installed Capital Cost Versus Flare Capacity (Fig. V-5)

- a. From Fig. IV-7 a 15-ft-diam enclosure flare is required.
- b. From Fig. B-2 a 15-ft enclosed ground flare installed capital cost is \$248,000.

2. Gross Annual Operating Cost (Fig. V-6)

- a. From Table V-2 the fixed cost is 29% of the installed cost:

$$248,000 \times 0.29 = \$71,900/\text{yr.}$$

- b. From Fig. IV-9 the natural gas used for the pilots is 100 scfh. From Table V-2 the cost for gas is \$2.00/thousand ft³:

$$100 \times 8760 \times \frac{2.00}{1000} = \$1750/\text{yr.}$$

c. Total Cost

Fixed	\$71,900
Gas	<u>1,750</u>
	\$73,650

3. Cost Effectiveness (Fig. V-7)

Cost effectiveness is the gross annual operating cost (B-2-c) divided by the annual waste-gas VOC destroyed at 99% efficiency.

a. Annual VOC destroyed = $20,000 \times 8760 \times 0.99 \times$ annual fraction of capacity operation.

b. Example at 1% of annual capacity operation:

$$\frac{73,650}{20,000 \times 8760 \times 0.99 \times 0.01} = \$0.042/\text{lb of VOC destroyed.}$$

4. Energy Effectiveness (Fig. V-8)

Energy effectiveness is the energy consumed for the pilots divided by the annual waste-gas VOC destroyed at 99% efficiency.

a. From B-2-b, natural gas consumed is 100 scfh; at 1000 Btu/ft³ amounts to 100,000 Btu/hr.

b. Example at 1% of annual capacity operation:

$$\frac{100,000 \times 8760}{20,000 \times 8760 \times 0.99 \times 0.01} = 505 \text{ Btu/lb of VOC destroyed}$$

C. FUEL-GAS SOURCE

This example is based on 850-lb/hr maximum waste-gas VOC emission rate.

1. Installed Capital Cost Versus Flare Capacity (Fig. V-9)

a. Figure IV-10 shows that a 2-in. fuel-gas transfer line is required.

b. Figure B-3 indicates that a 2-in. line installed capital cost is \$65,000.

2. Annual Operating Costs

a. From Table V-2 the fixed cost is 29% of the installed cost:

$$65,000 \times 0.29 = \$18,850/\text{yr.}$$

b. The electrical power consumption for compressing the gas must be calculated. For the model system of compressing from atmospheric pressure to 30 psig the adiabatic horsepower required was calculated to be 9.65

(Eq. 6-23 on page 6-16 of Perry's Chemical Engineers Handbook, 5th ed., McGraw-Hill, was used). Assuming an 85% electric motor efficiency and an 85% compressor efficiency and converting horsepower to kilowatts (1 hp = 0.746 kW), the electric power consumption rate is

$$9.65 \times \frac{1}{0.85} \times \frac{1}{0.85} \times 0.746 = 9.96 \text{ kW.}$$

From Table V-2 the electrical cost is \$0.03/kWh. For a 60% of annual capacity operation example the cost is

$$9.96 \times 8760 \times 0.03 \times 0.60 = \$1570/\text{yr.}$$

- c. From Table V-2 the fuel credit is \$2.00/million Btu. For waste gas with a net heating value of 19,600 Btu/lb this is equal to a credit of \$0.0392/lb of waste gas. For the example of 60% of annual capacity operation the credit is

$$850 \times 8760 \times 0.6 \times 0.0392 = \$175,130/\text{yr.}$$

d. Annual cost summary

Fixed	\$18,850	
Electrical	<u>1,570</u>	
Gross (Fig. V-10)	20,420	
Credit	<u>(175,130)</u>	
Net	(154,710)	Savings

3. Cost Effectiveness (Fig. V-11)

Cost effectiveness is the net annual operating cost (C-2-d) divided by the annual waste-gas VOC destroyed at 99.9% efficiency.

- a. Annual VOC destroyed = $850 \times 8760 \times 0.999 \times$ annual fraction of capacity operation.

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b. Example at 60% of annual capacity operation:

$$\frac{(154,710)}{850 \times 8760 \times 0.999 \times 0.60} = (\$0.0346)/\text{lb of VOC destroyed savings}$$